FLOWSHEET DEVELOPMENT AND COMPARISON FOR THE RECOVERY OF PRECIOUS METALS FROM CYANIDE LEACH SOLUTIONS

by

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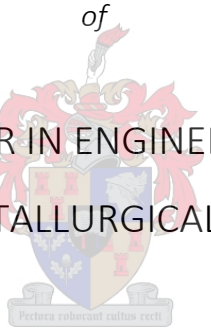

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Abstract

The Platreef ore deposit, situated in the Bushveld Igneous Complex, is one of the world's largest platinum group metal (PGM) resources. The mineralogy of this resource is, however, unique as it consists of complex PGM mineralization with mainly copper and nickel, at very low PGM grades. The PGMs are mainly present in the ore as slow floating refractory minerals resulting in marginal process economics when processing via traditional mill-float-smelt processes.

A new process is currently being investigated to extract PGMs from low grade Platreef ore and concentrate using a sequential heap leach process entailing heap bioleaching and high temperature cyanide leaching. The heap bioleach extracts the base metals in an acidic sulphate medium using a mixed culture of mesophiles and thermophiles. After heap bioleaching, the heap will be reclaimed, rinsed and restacked for high temperature cyanide leaching where the cyanide liquor is directly heated via solar energy in panels. Platinum, palladium and gold are extracted during the cyanide leaching stage and then recovered from the pregnant liquor either by adsorption onto activated carbon or ion exchange resins. Final metal recovery will proceed by techniques such as electrowinning and precipitation.

In this thesis, process options for the recovery of platinum group metals from cyanide solutions were identified with different flowsheet alternatives developed utilizing these options. Simulations were made for the different processing alternatives with the objective of finding the alternative flowsheet to maximise net present value. The various processing options were simulated, combining data from concurrent experimental studies and data reported in literature with kinetic adsorption models. This was combined with economic models to arrive at an optimum design for each flowsheet alternative.

Seven different processing alternatives for the recovery of platinum group metals from cyanide solutions were developed and investigated. These included two different activated carbon flowsheets as well as five different ion exchange resin flowsheets. The flowsheets differ in the elution procedures as well as the use of single or multiple resins. The well-known Merrill Crowe precipitation process was investigated but was found to yield unsatisfactory results.

Abstract

In each alternative, the cyanide solution is sent to a SART (sulphidization, acidification, recycling and thickening) plant to remove copper, nickel and zinc from solution prior to upgrading by means of adsorption onto activated carbon or ion exchange resins and subsequent elution. The platinum group metals are recovered from the eluate by precipitation using an autoclave, producing a solid product consisting of base and precious metals, while gold is recovered by electrowinning.

It was found that the overall performance of the resin-in-solution (RIS) flowsheets were superior to that of the carbon-in-solution (CIS) flowsheets, from an overall PGM recovery perspective and product grades. The superior adsorption kinetics and high selectivity of the resins for the PGMs resulted in excellent overall plant performances, with PGM extractions in excess of 97%. Gold extraction efficiencies with resins were found to be lower than those achieved in the CIS flowsheets, mainly due to the higher selectivity of the resins for the divalent platinum and palladium cyanide complexes and poor gold elution efficiencies. The gold concentrations in the feed streams to these processes were, however, very low, at only 8.5% of the total precious metal content. The overall precious metal recoveries of the RIS flowsheets were thus higher than the CIS flowsheets due to the superior PGM extractions.

From the cost analyses performed it was found that the RIS flowsheets requires lower initial capital costs, almost 28% lower than that require for the CIS flowsheets, while the operating cost requirements were found to be ±10% lower. This, combined with the high overall precious metal extractions, resulted in the RIS flowsheets to achieve higher net present values than those of the CIS flowsheets over an assumed project life of 15 years.

The optimum flowsheet proposed for the recovery of precious metals from cyanide leach solutions was a RIS flowsheet option that employed the Amberlite PWA 5 resin, capable of extracting platinum, palladium and gold from solution, with elution being performed with a zinc cyanide solution. This process option had the lowest capital and operating cost requirements while achieving similar overall precious metal recoveries as the other flowsheets. Economic analysis of this process yielded the highest net present value, with a 31% increase in the overall return on investment compared to the optimal CIS flowsheet.

Based on this, it was concluded that resin technology would be the best process option for recovering precious metals from cyanide leach solutions, however, additional research is required as the current level of process development is only at a concept phase.

Opsomming

Die Platrif-erts, geleë in die Bosveld Kompleks, is een van die wêreld se grootste platinum groep metaal (PGM) reserwes. Die mineralogie van hierdie reserwe is uniek en bestaan uit komplekse PGM mineralisasie met hoofsaaklik koper en nikkel, teen baie lae PGM inhoud. Die PGMe is hoofsaaklik teenwoordig in die erts as stadig drywende minerale en dit lei tot marginale ekonomiese uitsigte wanneer hierdie reserwe deur tradisionele metodes verwerk word.

Tans word 'n nuwe proses ondersoek om die PGMe vanuit lae graad Platrif-erts en konsentraat te ontgin deur gebruik te maak van 'n sekwensiële hooploogproses wat uit 'n bio-loog en hoë temperatuur sianied loog stappe bestaan. Die bio-loog stap is verantwoordelik vir die ontginning van die basis metale deur gebruik te maak van 'n suur sulfaat medium bestaande uit 'n gemengde kultuur van mesofiele en termofiele. Nadat die bio-hooploog stap voltooi is, word die hoop herwin, gewas en herpak vir die daaropvolgende hoë temperatuur sianied loog, waar die sianied oplossing direk verhit word deur die gebruik van son panele. Platinum, palladium en goud word tydens hierdie stap ontgin en kan dan herwin word vanuit die loog oplossing deur gebruik te maak van adsorpsie deur geaktiveerde koolstof of ioon-uitruilings harse. Finale metaal herwinning kan deur elektroplatering en presipitasie vermag word.

In hierdie tesis word verskeie proses opsies vir die ontginning en herwinning van PGMe vanuit sianied loog oplossings ondersoek waarna verskeie proses vloei diagram alternatiewe ontwikkel is met die doel om die opsie te vind wat die hoogste netto ekonomiese waarde sal oplewer. Die verskillende opsies is gesimuleer deur gebruik te maak van eksperimentele data gepubliseer in die literatuur en dit te kombineer met kinetiese adsorpsie modelle. Dit was dan gekombineer met ekonomiese modelle om 'n optimum ontwerp van elke proses te verkry.

Sewe verskillende proses vloei diagramme vir die herwinning en ontginning van PGMe vanuit sianied loog oplossings is ontwikkel en ondersoek. Hierdie het twee verkillende geaktiveerde koolstof prosesse en vyf verskillende ioon-uitruilings hars prosesse beslaan. Die opsies het verskil van eluerings metodes en adsorpsie medium. Die alombekende Merrill Crowe presipitasie proses is ook ondersoek, maar daar is gevind dat hierdie proses oneffektiewe resultate oplewer met betrekking tot die herwinning van die drie edel metale.

Opsomming

In elke alternatief word die sianied oplossing in ’n SART proses verwerk, waar die basis metale herwin word, gevolg deur die opgradering van die edel metale d.m.v. geaktiveerde koolstof adsorpsie of ekstraksie m.b.v. ioon-uitruilings harse, gevolg deur eluering. Die PGMe word dan herwin vanuit die eluerings oplossing deur termiese degradering van die metaal sianied komplekse, wat ’n hoë graad presipitaat lewer bestaande uit die basis en edel metale. Goud word herwin d.m.v. elektroplatering.

Daar is bevind dat die algehele verrigting van die ioon-uitruilings hars opsies beter was as die van die geaktiveerde koolstof opsies, beide van ’n algehele edel metaal herwinnings en produk suiwerheid perspektief. Die verhoogde adsorpsie kinetika en hoër PGM selektiwiteit van die harse het daartoe gelei dat uitstekende algehele PGM herwinning verkry is in hierdie opsies, meer as 97%. Goud ekstraksie deur die harse was laer as wat verkry was deur die geaktiveerde koolstof opsies, weens die hoër selektiwiteit van die harse vir die divalente platinum en palladium sianied komplekse en laer hars eluering effektiwiteit. Die goud konsentrasies in die voer strome na die prosesse was laag, en het sowat 8.5% van die totale edel metale uitgemaak, wat bygedra het tot die lae goud herwinning. Algeheel was die edel metaal herwinning van die hars prosesse beter as die van die koolstof prosesse a.g.v. die hoër PGM adsorpsie.

Koste evaluerings van die verskillende vloeidiagramme het getoon dat die hars opsies laer kapitaal kostes benodig, omtrent 28% minder as die koolstof opsies, terwyl bedryfskostes omtrent 10% minder was. Dit het bygedra tot die feit dat die ioon-uitruiling hars opsies ’n hoër algehele netto ekonomiese waarde oor ’n projek leeftyd van 15 jaar sal hê, aangesien de PGM ekstraksie, en dus die jaarlikse inkomste, ook hoër was.

Die algehele proses vloei diagram wat voorgestel is vir die herwinning van edel metale vanuit sianied loog oplossings is die hars opsie wat gebruik maak van die Amberlite PWA 5 hars, wat in staat is om platinum, palladium en goud terselfdertyd te absorbeer, gevolg deur die eluering van die hars deur die gebruik van ’n sink sianied oplossing. Hierdie proses het die laagste kapitaal en bedryfskostes getoon terwyl algehele PGM herwinning om en by dieselfde was as al die ander opsies. Hierdie proses sal verder ’n 31% verhoging in die opbrengs op belegging lewer in vergelyking met die optimum geaktiveerde koolstof opsie.

Die algehele gevolgtrekking is dat hars tegnologie die beter opsie sal wees vir die herwinning van edel metale vanuit sianied loog oplossings. Addisionele navorsing is dus nodig om resultate te verbeter aangesien hierdie studie slegs op ’n konsep fase benadering was.

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Chapter 1

Introduction

1.1 Project Background

The project was initiated due to the newly found interest in recovering precious metals from low grade platinum group metal (PGM) bearing material in the Bushveld Igneous Complex, and more specifically the Platreef ore body. The need to explore these low grade ore deposits has become a priority due to the depletion of the higher grade PGM resources, such as the Merensky and UG2 deposits, and the sharp increase in the market prices of precious metals over the last decade, with the platinum price rising from an average of $750/oz in 2003 to $2 000/oz in 2007. The platinum price has, however, dropped to about $1 400/oz in 2013, mainly due to the worldwide recession and instabilities in the South African mining sector, which is the largest producer of precious metals in the world.

The Platreef ore body, situated in the Limpopo province of South Africa, is a thick (between 20 - 100 meters) and complex mineral deposit as it shows marked variability in composition, mineralogy and hardness and consists of a complex assemblage of pyroxenites, serpentinites and calc-silicates (Gain & Mostert, 1982; White, 1994).

Platreef ore mineralization is unique and differentiates it from the Merensky and UG2 deposits. Base metal mineralisation and PGM concentrations are highly irregular in both value and distribution (Gain & Mostert, 1982; Schouwstra & Kinloch, 2000). The reef is predominantly a base metal (copper and nickel) and palladium resource by composition although the platinum value is still larger than the palladium value at current market prices.

The PGMs occur mainly as tellurides and arsenides, which is known to be slow floating when compared to the less abundant PGM sulphides (Schouwstra & Kinloch, 2000; Vermaak, 2005; Shamaila & O'Conner, 2008). The PGMs commonly occur enclosed within or on the grain boundaries of the base metal sulphides, such as pyrrhotite, pentlandite, chalcopyrite and pyrite, while there is also high association of PGMs with silicate minerals (Bryson, 2008).

As a result of this complex mineralisation and characteristics, recovery of PGMs via froth flotation is challenging as a significant amount of PGMs are lost to the tailings. Traditional MF2 (mill-float-mill-float) approaches to concentration of Platreef ores, such as Anglo American Platinum's Mogolakwena concentrator plant, has required higher than usual mass pulls to yield effective recoveries of the PGMs. This leads to a dilution in the overall grade and results in a poor quality concentrate with a high iron and sulphide content (Mogosetsi, 2006).

This, in turn, negatively impacts the downstream processes as very large investments in the subsequent smelting, gas handling and acid (or alternative sulphur capture) plant sections are required, which at the current market prices leads to marginal economics. Moreover, the area is remote, arid, has limited infrastructure and no significant skilled labour pool (Mwase *et al*., 2012*a;* 2012*b)*.

Due to these factors, an investigation into finding an alternative process has been initiated by Lonmin Plc, which could recover the precious metals from the Platreef resource and ultimately render the processing of this ore deposit economically attractive while placing a smaller burden on the water supply in the Olifants River Water Resource Development Project and reducing the electrical power demand and associated electrification delays.

1.2 Project Overview

Recently a new process has been proposed, shown schematically in Figure 1.1, to extract precious metals, mainly platinum, palladium and gold, from low grade Platreef ore or flotation concentrate using a sequential heap leach process, entailing heap bioleaching and high temperature cyanide heap leaching (Mwase *et al*., 2012*a*; 2012*b*).

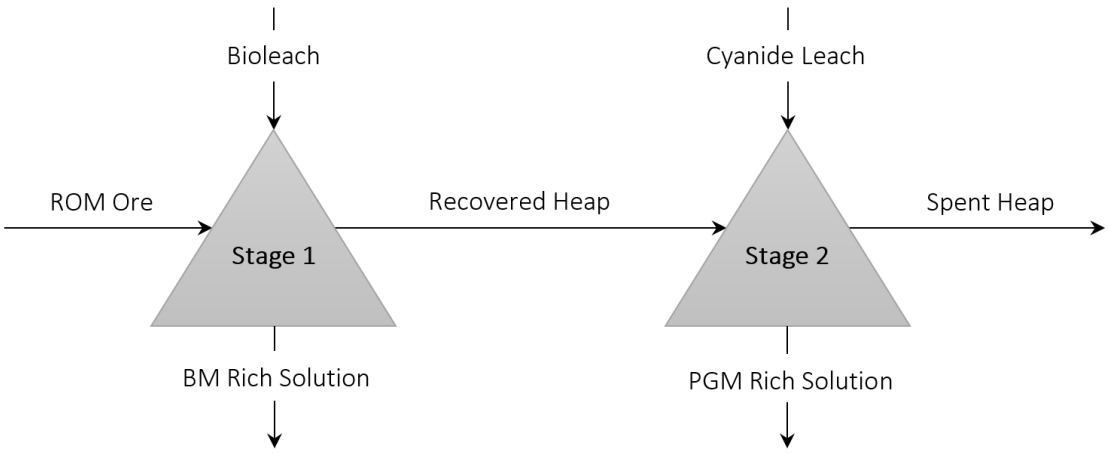

Figure 1.1*: Sequential heap leach for the recovery of base and precious metals from low grade PGM ore*

The heap bioleach is to extract the base metals in an acidic sulphate medium using a mixed culture of mesophiles and thermophiles. After heap bioleaching, the heap is to be reclaimed, rinsed and restacked for high temperature cyanide leaching, where the cyanide liquor is directly heated via solar energy with the use of solar panels. Platinum, palladium and gold are to be extracted during this cyanide leaching stage and a process is required in order to recover these metals from the pregnant liquor.

Currently there are no operating processes that recover platinum, palladium and gold from cyanide leach solutions as valuable high grade products that can be toll refined, or fed directly to a precious metal refinery. There are, however, well-established process options for the recovery of gold from dilute cyanide solutions.

In order to better understand the application of some of these established gold ore and other processing techniques to a low grade PGM ore, several experimental studies have been undertaken, which formed part of a major collaboration between the University of Cape Town and Stellenbosch University, to evaluate some of these processing steps individually for their use in processing Platreef ore and concentrate.

These included the sequential heap leaching of low grade Platreef ore and concentrate (Mwase, 2009), recovery of the base metals, leached during bio-heap leaching, by ion exchange resins (Liebenberg, 2012), adsorption of the PGMs in the cyanide leach liquor onto activated carbon (Mpinga, 2012) and ion exchange resins (Schoeman, 2012), the subsequent elution of the loaded carbon (Snyders, 2011) and resin (Schoeman, 2012) as well as the direct precipitation of the PGMs via Merrill Crowe zinc precipitation (Mpinga, 2012).

A study into the overall extraction and recovery process has, however, not been done. This project thus focuses on evaluating these downstream processing options for treating the pregnant liquor to recover the precious metals in a plant-wide optimisation study. In order to accomplish this goal and to assess which of these process options would yield the most cost effective method for recovering the precious metals from solution, different flowsheets incorporating these options and their alternatives need to be developed. These flowsheet alternatives can then be simulated using models and data adapted from gold recovery processes, together with data obtained during the aforementioned experimental studies, which if combined with economic models and subsequent optimization could be used to obtain an optimal design for each flowsheet.

Based on the results of these simulations and several other performance indicators, such as environmental factors, water usage and technical risk, it would be possible to propose the best process option for the recovery of precious metals from dilute cyanide solutions. This is the key issues that this project will address.

1.3 Project Objectives

The objectives of this study are to:

- Identify, from current available literature, process alternatives for recovering precious metals from cyanide leach solutions,
- Investigate all these alternatives and possible variations thereof, and develop different flowsheets based on these variations,
- Simulate these flowsheets using data from con-current experimental studies and models from literature,
- Evaluate these flowsheets based on circuit design and precious metal recovery,
- Evaluate the profitability of each of these flowsheets based on economic estimates,
- Perform economic optimization for these flowsheets based on maximising the net present value (NPV) and select the best flowsheet based on this objective function.

1.4 Research Questions

The main research questions to be answered in this project are:

- What are all the process alternatives available for recovering precious metals from cyanide leach solutions?
- What are all the variations of these alternatives?
- What are all the challenges associated with these process alternatives?
- Which flowsheet would yield an optimal design based on process economics?
- Which flowsheet is the best process option for recovering precious metals from cyanide leach solutions based on factors such as capital and operating costs, product quality and project profitability?

1.5 Scope Limitations

Due to the specific nature of this project and to adhere to the requirements of the financial sponsor of this project, Lonmin Plc, several limitations were placed on the overall project scope. These limitations are listed below:

- The aim is to specifically look at the downstream processing options for treating cyanide leach solutions with specific reference to carbon adsorption and ion-exchange technologies. Thus the Platsol (Dreisinger *et al*., 2009), Kell (Liddell & Adams, 2012) and other processes were not considered.
- The aim is to investigate a process incorporating carbon adsorption or ion exchange resins that can produce a PGM precipitate suitable for the direct feed to a PMR or a product that can be toll refined. Thus no alternatives for the production of an intermediate product, such as bulk sulphide precipitation, were investigated.
- Cost analysis for the associated mining, leaching and refining processes were not performed as these were already performed in-house by Lonmin Plc. The aim is to investigate and compare several downstream flowsheet options for treating the cyanide leach solution and not to provide a business model for the overall mining operation.
- It is thus not the aim of this study to compare the overall process for recovering PGMs from the Platreef ore with similar PGM projects, PGM ore reserves or potential concentrator plants. The justification of treating the Platreef with a cyanide heap leach solution is thus not part of the project scope.

1.6 Thesis Structure

The work in this study is presented as follows:

Chapter 2: Platreef Processing Options

A review of research conducted on complex ore treatment processes and process options for the treatment of cyanide leach solutions are presented in this chapter. This chapter mainly focuses on treatment options used for the recovery of gold from dilute cyanide leach solutions.

Chapter 3: Application of gold processing techniques to low grade PGM ore

This chapter primarily focuses on the findings of several con-current experimental studies performed, based on the application of several of the treatment options discussed in chapter 2, for recovering precious metals from cyanide leach solutions.

Chapter 4: Flowsheet Development

In this chapter all the possible process options are discussed based on their applicability for recovering precious metals from cyanide leach solution. Different flowsheets that will be used to ultimately propose the best treatment option are then identified and discussed.

Chapter 5: Basis of Design

The main assumptions made and design conditions used in order to allow for the simulation of each of the different unit operations, and hence the different overall flowsheets, are presented in this chapter.

Chapter6: Process Economics

The methods used in estimating the different economic parameters for each flowsheet are presented in this chapter. This includes capital cost estimations (CAPEX), operating cost estimations (OPEX), net present value (NPV) and present value ratio (PVR) calculations as well as Monte Carlo simulations.

Chapter 7: Adsorption Modelling and Optimization

Adsorption experiments were conducted to determine the kinetic constants required to ultimately model, size and optimise the adsorption sections of the different flowsheets. The results of these experiments are presented in this chapter as well as a review of different adsorption kinetic models. A comprehensive review of the selected model is presented, while a discussion on the subsequent modelling and optimization methods used rounds of this chapter.

Chapter 8: Simulation Results and Flowsheet Comparison

The optimized simulation results are presented in this chapter followed by a detailed comparison between the different flowsheets based on their respective product quality, process economics and other performance indicator. Finally a single flowsheet option is proposed that would achieve a high overall PGM recovery at favourable economics.

Chapter 9: Conclusions and Recommendations

The main conclusions and recommendation drawn from the comparisons of the different flowsheets as well as the optimum flowsheet proposed are presented in this chapter. Recommendations as to further improve the accuracy of the results of this study as well as possible future work are also discussed.

Chapter 2

Platreef Processing Options

The following chapter is a review of the literature that was found to be applicable to this research project. A basic overview of platinum group metals and the Platreef ore is given followed by a review of ore types and the challenges that such ores present in recovering the value material. Current technologies and process options available for treating these ores are discussed based on the premise/hypothesis that adapting processes currently used for recovery of gold from dilute cyanide leach solutions would give a process suitable for the recovery of platinum, palladium and gold. This chapter mainly focuses on determining which of the processing options used for the recovery of gold from leach solutions would be appropriate for its application to the Platreef ore body.

2.1 Platinum Group Metals Overview

The family of platinum group metals (PGMs) comprises platinum, palladium, rhodium, iridium, osmium, and ruthenium. The PGMs have high melting points and stability at high temperatures. They are also extremely resistant to corrosion and have exceptional oxidation and reduction abilities (Renner *et al*., 2001).

Platinum group metals are of significant value as they are utilized in a number of industrial processes, technologies and commercial applications. The unique properties of platinum (chemical and physical) make it an excellent raw material, catalyst or ingredient in other manufacturing processes (Renner *et al*., 2001).

Platinum is used in many different applications. These include the coinage and investment industry, jewellery market, medical apparatus production, catalyst and fuel cell production as well as small electric parts used in sensors and electrical technologies (Renner *et al*., 2001; Jollie, 2009; Cowley, 2013).

Around 20% of the total platinum production is reserved for the jewellery industry with markets rapidly expanding in the far and Middle East. Automotive catalytic converter applications are, however, the largest current users of PGMs. These devices are used to treat exhaust gases, reducing pollutants and with higher clean air standards being implemented across the world, the demand for platinum is constantly rising, although some manufacturers have begun to turn to lower-cost palladium for this function (Renner *et al*., 2001; Jollie, 2009).

Both platinum and palladium continue to be very rare and much of the world's platinum production comes from the small Western Bushveld Complex, in South Africa, which accounts for over 70% of the world's supply (Renner *et al*., 2001; Cowley, 2013). The need to exploit this low grade resource has become more important as the worldwide demand of platinum and palladium increased by 5% and 12.5% respectively over the last two years, while the worldwide supply has dropped by 11.5% and 12.6% respectively (Cowley, 2013).

Current available process options, however, result in uneconomic exploration of the lower grade and more complex ore deposits, such as the Platreef which is part of the Bushveld Igneous Complex. The need for extraction processes to be developed and improved are thus becoming more and more important as new technologies are required to supply the world demand, while competition from PGM recycling further threatens the mining sector.

This project is thus a step in this direction as the aim is to propose a process for the recovery of PGMs from cyanide leach solutions that meets both economic cost requirements and recovery efficiencies. This will enable the extraction and recovery of PGMs from low grade ores, which could result in previously uneconomic ore deposits to be treated, if successful.

2.2 The Platreef Ore Deposit

The Platreef ore body is situated in the lower critical zone of the northern limb of the Bushveld Igneous Complex and consists of a variety of minerals bearing the valuable precious metals, mainly platinum and palladium. The resource is predominantly a copper-nickel resource (0.2 - 0.3% copper and 0.15 - 0.2% nickel) and is low in overall PGM grades (2 - 5 g/t) but still hosts a large portion of the worldwide PGM reserves (Cramer, 2001). The Platreef, however, remains poorly exploited at present due to the complex PGM mineralization, reef thickness and the multiple challenges that this presents, especially for resources that cannot be mined by open pit methods.

As mentioned before, the Platreef consists of a complex assemblage of pyroxenites, serpentinites and calc-silicates, with the PGMs mainly present in the form of the slow floating telluride and arsenide minerals (Gain & Mostert, 1982; White, 1994; Schouwstra & Kinloch, 2000; Vermaak, 2005; Shamaila & O'Conner, 2008). The PGMs are further found enclosed or on the grain boundaries of various base metal minerals, mainly sulphides such as pyrrhotite, pentlandite, chalcopyrite and pyrite. The Platreef mineralogy is further unique in that that the base metal mineralization and PGM concentrations vary significantly throughout the ore body both in value and distribution (Schouwstra & Kinloch, 2000).

The main platinum group and base metal minerals commonly present in the Platreef are presented in Table 2.1 (Schouwstra & Kinloch 2000; Seymour and O'Farrelly 2001; Crundwell *et al*., 2011). The next sections focus on typical flowsheet options for the recovery of most of these metals by the use of mainly cyanidation techniques.

Table 2.1: *Metal Groups and its constituents*

Platinum Group Minerals		*Base Metal Sulphides*		*Others*
Cooperite	(PtS)	Pyrrhotite	(FeS)	Silicates
Moncheite	$(PtTe_2)$	Pentlandite	$[(Fe,Ni)_9S_8]$	Chromite
Laurite	$[(Ru,Os,Ir)S]$	Chalcopyrite	$(CuFeS_2)$	
Ferroplatinum	(Pt_3Fe)	Pyrite	(FeS_2)	
Sperrylite	$(PtAs_2)$	Millerite	(NiS)	
Braggite	$[(Pt,Pd)NiS]$	Cubanite	(Cu_5FeS_4)	

2.3 Process Flowsheet Selection

The competing technologies and their respective advantages and economics for the treatment of gold and precious metal ores have become a major focus point within the mineral processing industry. The reason for this has been the increasing refractory and complex nature of the ore bodies that are treated, while the possible extraction of gold and precious metals from previously uneconomic ore bodies and tailings streams has prompted the mineral processing industry to develop techniques that could yield economically favourable results. These techniques are hugely dependent on the type of ore to be treated.

La Brooy *et al.* (1994) provided a framework for gold ore bodies to be classified into three different categories that determine the processes that could be used to treat them economically. These categories are listed below:

- **Free-milling Ores:** This type of ore generally yields recoveries in excess of 90% when treated with conventional cyanidation processes.
- **Complex Ores:** This type of ore gives only economic recoveries when treated with significantly higher chemical additions such as cyanide, oxygen and carbon. They typically consist of base metal mineralization which increases cyanide consumption and introduces problems in the operation of a carbon adsorption circuit.
- **Refractory Ores:** This type of ore gives uneconomic recoveries under conventional cyanidation treatments and is very difficult to treat. Usually some sort of pre-treatment prior to cyanidation is required.

The Bushveld Igneous Complex, and more specifically the Platreef ore body, is classified as a complex ore as it consists of extensive base metal mineralization and very low grades of valuable precious metals (Gain & Mostert, 1982; White, 1994; Schouwstra & Kinloch, 2000; Vermaak, 2005; Shamaila & O'Conner, 2008). The different process alternatives for such an ore and the challenges and selection issues associated with it are discussed below.

2.4 Complex Ore Processing Options

Complex ores, as mentioned above, are classified as being between free-milling and refractory ores. They are thus neither the easiest nor the most challenging ore bodies to treat and extensive base-metal mineralization exists which introduces new challenges. Complex ores treated via cyanidation methods consume vast amounts of cyanide and high oxygen levels may further be required during carbon adsorption. Complex ores may also be preg-robbing, thus containing natural carbon that adsorbs the gold prior to the carbon adsorption stage, which could greatly affect the efficiency of such a process (Fleming, 1992; Lunt & Weeks, 2005).

The discussion that follows focuses on relevant cyanidation treatments and the challenges that complex ores may provide, giving insight into the challenges for treating the Platreef ore body.

2.4.1 High-copper Ores

Ores with a high copper content are almost always treated via flotation and smelting with the gold being recovered in the refinery (Adams, 1999). The complexity and low grades of these ores, such as Platreef which is predominantly a copper and nickel resource, do not allow for such techniques to be used economically. Cyanidation techniques, such as heap or pressure leaching, are typically employed for extracting the gold from these ores, with the gold recovered in a carbon adsorption or Merrill Crowe circuit (Lunt & Weeks, 2005).

Unfortunately, copper minerals are extensively soluble in cyanide and this introduces many challenges. Typical copper complexes found in cyanide liquors are $Cu(CN)_2^-$, $Cu(CN)_3^{2-}$ and $Cu(CN)_4^{3-}$. The amounts of each present in the liquor are dependent on the pH, free cyanide concentration and the copper concentration. The formation of the low $Cu{:}CN$ mole ratio species ($Cu(CN)_2^-$) is favoured to minimize the cyanide consumption, a major cost driver during cyanidation (Lunt & Weeks, 2005)

These copper complexes are readily adsorbed onto activated carbon and directly influence the gold adsorption (Nicol *et al*., 1984*a*, Fleming, 1992; Marsden & House, 2006; Dai & Breuer, 2009; Dai *et al*., 2010; Mpinga, 2012). One way of rejecting the copper is to use a cold cyanide solution during elution to preferentially strip the copper from the carbon prior to the elution of gold (Marsden & House, 2006; Snyders, 2011).

For ores containing high amounts of soluble copper, however, more specific flowsheet measures are required, which can be divided into two groups: processes that aim to minimize cyanide consumption and processes that aim to recovering the cyanide/copper from solution.

For the minimization of cyanide consumption pre-treatment by acid leaching or vat-leaching, with the use of ammoniacal cyanide leachant, is required, that dissolves the more soluble copper oxides and sulphides prior to the cyanide leach (Lunt & Weeks, 2005). Research has also been conducted on a bio-leach to extract base metals prior to the cyanide leach (Mwase, 2009; Mwase *et al*., 2012*a*, 2012*b*)

Copper and cyanide recovery processes include the AVR (acidification, volatilization, regeneration) process, introduced at the Beaconsfield gold project in Tanzania (Kitney, 1998) and the relatively new SART (sulphidization, acidification, recycling, thickening) process, introduced at the Tefler project in Australia (Barter *et al*., 2000).

In the SART process the copper is removed from solution prior to the adsorption stage with cyanide recovered for re-use in the leaching step. This process is believed to greatly increase the efficiency of a carbon adsorption circuit and could be applied directly after heap-leaching (Ford *et al*., 2008).

In the case of the Platreef ore, which contain high amounts of soluble copper, a copper minimization or reduction step will be required to negate the negative impacts that the copper may have on the downstream processes. This could be accomplished by employing either an acid or bio leaching step, extracting the copper and other base metals, or the treatment of the cyanide leach solution in an AVR/SART plant.

2.4.2 Oxygen Consuming Ores

Some ores, particularly those consisting of high amounts of sulphide minerals (especially stibnite and pyrrhotite), such as the Platreef ore, consume oxygen during the cyanide leaching and carbon adsorption stages (Lunt & Weeks, 2005). Under some conditions the oxygen requirement can be very high, and if not supplied, could drastically reduce the performance in these stages, as a drop in the dissolved oxygen levels in the pulp or solution could reduce the rate of gold dissolution and ultimately the completion thereof. If such an ore is processed, oxygen injection is required and can be best utilized if an air-sparged agitation stage is included ahead of the carbon adsorption stage (Lunt & Weeks, 2005)

2.4.3 High-silver ores

The flowsheet selection options for the extraction and recovery of gold ores containing silver is based on the silver content of the ore, the total output of precious metals and the solid/liquid separation characteristics of the leached pulp (Allen *et al*., 1998; Lunt & Weeks, 2005).

A typical gold ore extraction and recovery circuit would be a leach stage followed by carbon adsorption and elution with the gold being recovered by electrowinning (Fleming, 1992; Stange, 1999; Lunt & Weeks, 2005; Marsden & House, 2006). This process route, however, becomes unfavourable when the amount of silver within the ore reaches high levels, in excess of 10:1 (Marsden & Fuerstenau, 1993). Considerations should then be made as to the process route to be taken, either a zinc precipitation (Merrill Crowe) circuit in combination with carbon adsorption or if the silver throughput is too high, a full zinc precipitation circuit.

Allen *et al*. (1998) examined the alternative processing routes for the Rawas ores on their silver/gold ratios. These findings have been developed into guidelines for the process selection for treating ores containing silver and precious metals based on their ratio and are given below:

- Low - Typical CIL/CIP - elution - electrowinning
- Moderate - Consider CIL/CIP - elution - zinc precipitation
- High - Consider full operation of a Merrill Crowe plant

There are thus several options available for the treatment of gold ores containing high amounts of silver. This is an important consideration if the PGMs, such as platinum and palladium, were found to behave similarly to silver in the above mentioned processes, as the PGM to gold ratio of the Platreef ore is very high. Zinc precipitation may further be an attractive option as the costly counter-current decantation (CCD) step, or solid liquid separation, that is necessary for treating slurries or pulp is not required as heap leaching of the Platreef ore is proposed.

2.4.4 Recovery of PGMs from Leach Solutions

Milbourne *et al.* (2003) has described several concepts/methods for the recovery of PGMs from leach solutions. Dissolved PGMs introduce difficulties during recovery as they are strong oxidants and are easily reduced to their respective metals. They can also be easily hydrolysed and thus tend to precipitate at unwanted locations which can pose many difficulties in the design and development of a robust and controllable process.

Salt precipitation can be used as there are numerous compounds of these metals used in refining processes. Platinum and palladium can be precipitated with the use of ammonia or sulphide under the right circumstances but the low concentrations of typical leach solutions may deem the addition of these salts uneconomical.

As mentioned earlier, PGMs are strong oxidants and can be relatively easily reduced to their respective metals with the use of sulphides, ferrous iron, zinc metal, alkenes or oxalic acid. The concentrations of these reduced metals would be low and a carrier, like a base metal sulphide, may be required to assist in the recovery of the precious metals from solution. This would, however, reduce the purity of the desired product leading to a decrease in product quality.

Adsorption by ion exchange or activated carbon is well known for the concentrating of dilute leach solutions. An adsorbent with a high selectivity affinity for the desired ions is required to produce favourable results. This may include resins (chelating, strong base or reducing) or activated carbon. The resin or carbon can then be eluted resulting in a more concentrated solution that could make the recovery process easier.

Solvent extraction can also be utilized. It is, however, not a desirable approach when treating dilute solutions typically obtained during heap or concentrate leaching, as the cost of reagent and metal losses can easily match the value of the metals in the pregnant solution. It could, however, be applied to concentrated solutions more effectively, such as that produced during carbon elution, as it reduces the amount of reagent required.

2.4.5 Summary of Complex Ores Process Options

From the preceding discussion it is clear that there is no fixed processing route for the treatment of gold/PGM ores as the nature of the ore and its mineralogy greatly affect the operation and performance of each step during the recovery process. Various processing alternatives thus exist based on the type of ore.

Typically some sort of a cyanidation approach is followed. The processing steps in such a process, as identified in the preceding discussions, are typically a leaching stage (heap/tank), an absorbent/concentrating stage (SART/carbon adsorption/ion exchange resins) and a recovery stage (zinc precipitation/electrowinning). Each one of these processes has its own advantages and disadvantages and reasons for use based on their respective application to a certain ore. These process alternatives are reviewed in more detail in the next section.

2.5 Process Options for the Recovery of PGMs from Low Grade Ores

The process under development for the recovery of precious metals from cyanide leach solutions investigated experimentally a number of processing steps adapted from well-known gold recovery processes. In this thesis, results from these experimental investigations are combined with models to enable overall analysis of various flowsheet options to be made. These include cyanide heap leaching, the newly developed SART process, carbon adsorption, ion-exchange technology and the widely used Merrill Crowe zinc cementation process. The background and operation of these processes are reviewed in the flowing sections.

2.5.1 Cyanide Heap Leaching

The first metallurgical step in modern gold plants today is some sort of leaching process. In the process proposed for the recovery of PGMs and gold from the Platreef ore, cyanide leaching of the precious metals will be used on the heap following base metal leaching (Mwase *et al*., 2012*b*). This processing step is discussed below.

2.5.1.1 Overview

Cyanide heap leaching is a hydrometallurgical development for exploiting low-grade ores, mine waste material, or deposits that are too small to justify the construction of a typical concentrator plant (Heinen *et al*., 1978; Marsden & House, 2006). This technique is used to leach valuable minerals that are present in very low concentrations (ppm levels) to enable downstream processing of the solution that could result in an economically viable process (Heinen *et al*., 1978; Fleming, 1992; Marsden & House, 2006).

Cyanide heap leaching has been used for many years for the recovery of copper from oxidized ores but the first commercial scale application of this process for the recovery of gold was at the Carlin mine in 1971 (Heinen *et al*., 1978; Marsden & House, 2006). It was based on development work by the United Sates Bureau of Mines at Salt Lake City in the late 1960's and at Carlin mine itself from 1968 (Marsden & House, 2006).

This process has opened up vast reserves of low-grade ores that was previously deemed uneconomical to exploit and has probably contributed more than any other process to the uprising of gold production in the early 1980's (Fleming, 1992).

Leaching of low-grade precious metal ores with the use of alkaline cyanide solutions yields a metal bearing solution containing relatively low concentrations of gold, silver, copper, nickel, zinc and other metal cyanide complexes. This solution can then further be treated in subsequent downstream processes such as carbon adsorption, zinc cementation or ion exchange processes to recover the valuable material from solution (Hampton, 2002).

Cyanide heap leaching used for the extraction of gold from low grade ores is a very low cost (1/3 of the capital cost of an agitated leach plant) and flexible process option as it is further well-suited for both small (150 - 300 t/m) and large scale operations (300 000 t/m) (Kappes, 2005). Recoveries similar to that obtained in conventional milling and flotation circuits can be achieved resulting in a higher return on investment (Fleming, 1992; Kappes, 2005).

The process under development is designed for treating around 240 000 t/m and if cyanide heap leaching is found to be effective in recovering the PGMs from Platreef ore, treatment of the Platreef ore body could become more economically attractive. Promising results have been reported recently for the extraction of PGMs by cyanide heap leaching (Mwase, 2009; Mwase *et al*., 2012*a*, 2012*b*).

2.5.1.2 Process Description

Cyanide heap leaching involves the stacking of the metal-bearing ore onto sloped impermeable pads, irrigating the ore with a dilute cyanide solution allowing the cyanide to percolate through the heap dissolving the sought after metals, and collecting the pregnant leached solution for recovery of the value material in the downstream processes (Heinen *et al*., 1978; Fleming, 1992; Kappes, 2005).

Cyanide heap leaching are used commercially in two methods, short-term leaching of crushed ore (or low grade flotation concentrate) and long-term leaching of run-of-mine ore (Heinen *et al*., 1978).

Short-term leaching: The ore is crushed and stacked onto permanent pads at a height of between 1 - 3 meters. The ore is leached by sprinkling dilute cyanide solution onto the top of the heap and allowing the solution to pass through the heap dissolving the value material as it comes in contact with it. The solution, containing the dissolved gold, flows down the sloped watertight pads and is collected and transferred to a storage tank or pond.

The stacked ore is crushed to a size that gives good liberation and exposure to the cyanide solution, typically between 5 - 20 mm. This allows for the cyanide solution to quickly percolate through the ore and dissolve the value material so that the cycle can be repeated every 7 - 30 days (Heinen *et al*., 1978). The cycle is typically complete when the value material concentration drops below an economic cut-off value of 0.05 - 0.1 mg/L (Fleming, 1992). After the cycle has been completed the ore is removed and replaced with newly crushed ore. These types of operations can go up to about 10 000 t/d (Heinen *et al*., 1978).

Long-term leaching: In long-term leaching the ore being leached is normally uncrushed, porous, sub-mill-grade material or otherwise stated run-of-mine ore coming straight from the blasting operations underground. The ore particle size ranges from between 5 - 15 cm. The ore is stacked in a truncated pyramid shape between 6 - 10 meters in height ranging from 10 000 - 2 000 000 tons. The height of the heap is determined by factors such as ore permeability, cyanide strength and sufficient dissolved oxygen levels (Heinen *et al*., 1978). The leach cycle for such an operation can be several weeks, months or even years depending on the type of ore. (Fleming, 1992).

Cyanidation of the Platreef ore has been reported to be effective in recovering the PGMs from the ore (Mwase *et al*., 2012*b, 2014*). Platinum dissolution is, however, slow and has not reached a stationary value or completion even after more than 300 days of leaching. The process under development is further designed for treating about 240 000 t/m, which would result in significant comminution costs if short term leaching is required, due to the fine grind size required to ensure effective leachate flow characteristics. Taking this into consideration long term leaching would possibly be the method employed.

Several factors affect the efficiency of a heap leaching operation (Kappes, 2005). These include:

- Type of ore - The mineralization of the gold greatly influences the dissolution of the gold by cyanide
- Climate – The ideal heap-leach location is a temperate semi-arid desert location. It can, however, be applied successfully in different climates
- Heap permeability and flow efficiency – The stacking method greatly affects the wash efficiency of the ore. Ideal plug flow is desired
- Cyanide application rate, strength and leach time – All these factors contribute to the efficiency of gold dissolution

Cyanide heap leaching operations have a typical cyanide application rate of 6 - 60 L/h.m^2. The concentration of the cyanide solution for typical gold leaching operations is in the range of 200 - 600 mg/L $NaCN$. The pH within the heap is kept at about 10 with the application of lime or caustic soda (Heinen *et al*., 1978; Kappes, 2005; Marsden & House, 2006).

Gold is typically present in the ore in the elemental form or as an alloy with silver (Heinen *et al*., 1978; Hampton, 2002). PGMs in the Platreef ore are typically found enclosed in or on the grain boundaries of base metal sulphides and silicates, such as pyrrhotite, pentlandite, chalcopyrite and pyrite, as refractory tellurides and arsenides or as part of intermetallic ferroalloys (Gain & Mostert, 1982; White, 1994; Schouwstra & Kinloch, 2000).

The basic principle of cyanide heap leaching is that the weak alkaline cyanide solution has a dissolving action on the gold. This dissolution of gold by cyanide is represented by the following overall reaction, known as Elsner's equation (Heinen *et al*., 1978; Hampton, 2002):

$$4Au + 8CN^- + O_2 + 2H_2O \rightarrow 4Au(CN)_2^- + 4OH^-$$

The gold or aurocyanide complex ($Au(CN)_2^-$) formed during leaching is the predominant species formed and stays stable at very low pH and in the absence of free cyanide, making it a very stable complex to work with in downstream processes (Hampton, 2002).

The dissolution reactions of platinum and palladium can occur at ambient conditions, as in the case of gold dissolution (McInnes *et al*., 1994) and forms the stable $Pt(CN)_4^{2-}$ and $Pd(CN)_4^{2-}$ complexes. Chen and Huang (2006) have reported that the PGM dissolution reactions also follow Elsner's equation, and are given below:

$$2Pt + 8CN^- + O_2 + 2H_2O \rightarrow 2Pt(CN)_4^{2-} + 4OH^-$$

$$2Pd + 8CN^- + O_2 + 2H_2O \rightarrow 2Pd(CN)_4^{2-} + 4OH^-$$

Cyanidation of the PGMs at ambient conditions is, however, very slow due to poor kinetics and low extraction efficiencies of 15% platinum and 44% palladium have been reported (McInnes *et al*., 1994; Torres & Costa, 1997). Higher temperatures (100 - 125°C) and elevated pressures (1.5 - 1.8 MPa) are required to achieve effective extractions (Chen & Huang, 2006). Extractions in the order of 60 - 96% platinum and 70 - 98% palladium have been reported at these conditions (Torres & Costa, 1997; Chen & Huang, 2006).

During the cyanidation of these gold and PGM bearing ores several base metals are co-dissolved, with copper being the most dominant and most problematic in downstream processes. The dissolution reaction for copper sulphide (chalcocite) is given below (Hampton, 2002; Marsden & House, 2006).

$$Cu_2S + 7CN^- + O_2 + 2H_2O \rightarrow 2Cu(CN)_3^{2-} + CNS^- + 4(OH)$$

The most important complexes that are formed by cyanidation during heap leaching and which are essential in recovery of the value material are listed below:

- $Au(CN)_2^-$
- $Pt(CN)_4^{2-}$
- $Pd(CN)_4^{2-}$
- $Ag(CN)_2^-$
- $Hg(CN)_4^{2-}$
- $Cu(CN)_3^{2-}$
- $Ni(CN)_4^{2-}$
- $Co(CN)_6^{3-}$
- $Zn(CN)_4^{2-}$
- $Fe(CN)_6^{4-}$

2.5.2 S.A.R.T. Process

The Platreef ore body is predominantly a copper nickel resource, and consists of extensive precious metal mineralization with these base metals (Schouwstra & Kinloch, 2000). These base metals are readily leached during cyanide heap leaching (Fleming, 1992; Hampton, 2002; Mwase, 2009; Mwase *et al*., 2012*a*, 2012*b*) and creates metallurgical challenges by consuming the cyanide reagent, which negatively impacts PGM and gold yields (Tran *et al*., 1997; Botz *et a*l., 2011; Ford *et al*., 2008), while also adversely affecting the downstream recovery processes, such as carbon adsorption (Fleming, 1992, Fisher & LaBrooy, 1997; Marsden & House, 2006) and electrowinning (Steyn & Sandenberg, 2004). Several of these impacts are listed below:

- Increase in cyanide consumption during heap leaching
- Reduces the ability of the carbon/resin circuit to effectively adsorb the precious metals
- Increases the elution cycle, as it blocks carbon pores
- Introduces handling difficulties during gold electrowinning

The result of this high copper content associated with precious metal deposits is that some of these deposits are not exploited due to the metallurgical challenges, high processing costs and environmental issues associated with treating such ores. This led to the development of the so-called SART process for the removal of copper from leached solutions (Barter *et al*., 2000; Botz *et al*., 2011; Ford *et al*., 2008).

2.5.2.1 Overview

The SART process was jointly developed into a feasible process by SGS Lakefield Research and Tech Corporation in 1997 and was first introduced at the Tefler mine in Australia (Barter *et al*., 2000). It was developed as a potential solution for gold projects in which the presence of soluble base metals, that consume cyanide, greatly affected the gold recoveries and economics (Barter *et al*., 2000; Botz *et al*., 2011; Ford *et al*., 2008).

The process name, SART, is derived from an acronym describing the different unit operations in the process flowsheet:

- Sulphidization – The addition of sodium sulphide (Na_2S) or sodium hydrosulphide (NaHS) allowing the precipitation of Cu_2S
- Acidification – The addition of sulphuric acid to lower the pH to allow precipitation of copper to occur
- Recycling – Recycling of precipitate to allow the formation of larger crystals to ensure better settling performance
- Thickening – The thickening of the precipitate to be sold as a valuable by-product

In a typical SART process the cyanide associated with the soluble base metal complexes is released by the addition of a dosing agent, sodium hydrosulphide ($NaHS$), which is responsible for the precipitation of mainly copper, zinc and nickel under weak acidic conditions. This allows for the recycling of the cyanide back to the heap leaching stage as free cyanide for gold recovery. The recycling of cyanide allows for operation at higher cyanide levels, increased leaching efficiency and minimizes the effects of copper in downstream processes (adsorption and electrowinning circuits) (Barter *et al*., 2000; Botz *et al*., 2011; Ford *et al*., 2008).

Up to 99% of the copper can be recovered as a valuable high-grade (~70%) copper sulphide (Cu_2S) by-product (Botz *et al*., 2011). The potential presence of precious metals in this solid increases the value of the sulphide precipitate. Although some cyanide loss is found, 95% of the cyanide is regenerated as free cyanide used to increase the overall performance of the gold process (Botz *et al*., 2011).

2.5.2.2 Process Description

As mentioned above the SART process was designed to control the concentration of soluble base metal complexes in cyanide leached solutions and to free up additional cyanide for recycle and reuse. This process may be applied to the pregnant leach solution prior to carbon and resin adsorption or on the barren solution after adsorption to recover residual base metal losses and regenerate cyanide for re-use in the heap leach step (Ford *et al*., 2008).

The process has a distinct advantage if applied prior to adsorption as it eliminates the negative effects that base metal complexes have on the adsorption efficiency of the carbon and further eliminates the need for a cold cyanide strip to remove copper prior to gold during elution (Barter *et al*., 2000; Botz *et al*., 2011; Ford *et al*., 2008).

In a typical SART process, schematically shown in Figure 2.1, the pregnant leach solution originating from either a heap leach or tank leach stage is fed to the base metal precipitation reactor where the solution is acidified by the addition of sulphuric acid (pH 5 or below). At this low pH the addition of sulphide ions, in the form of sodium hydrosulphide ($NaHS$), results in the breakup of the weak acid dissociable base metal cyanide complexes and precipitation of copper, cobalt, nickel, zinc and minimum amounts of precious metals as metal sulphides, while converting cyanide to HCN. The overall acidification and sulphidization of copper is given by the following reaction (Botz *et al*., 2011):

$$2Cu(CN)_3^{2-} + 3H_2SO_4 + S^{2-} \rightarrow Cu_2S_{(s)} + 6HCN_{(aq)} + 3SO_4^{2-}$$

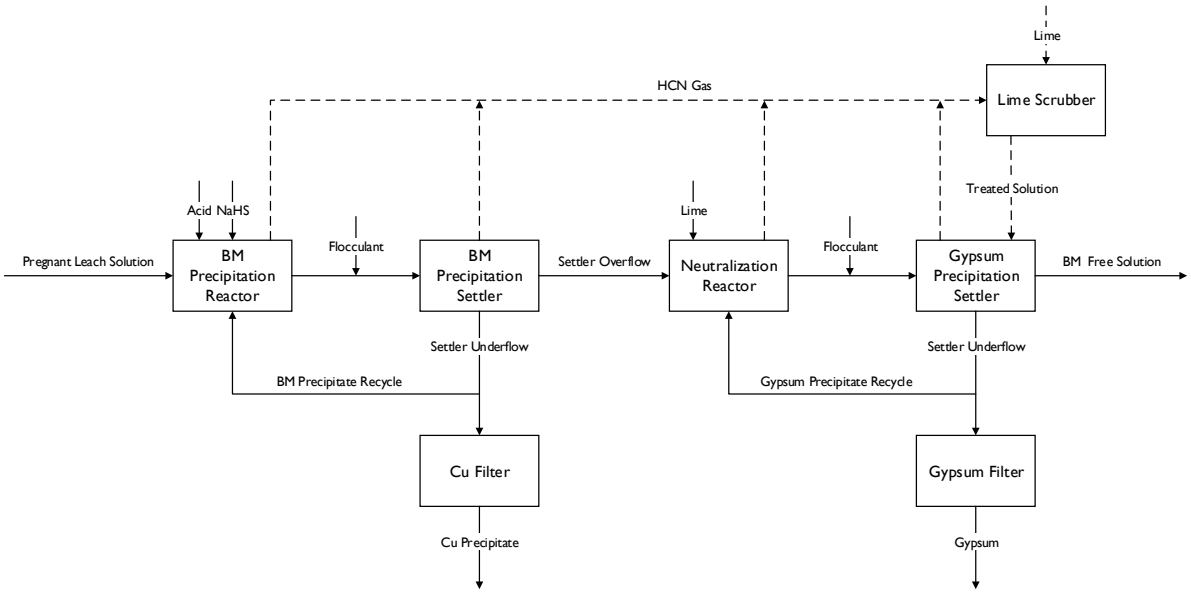

***Figure 2.1**: SART Block Flow Bigram*

The settling of these precipitates occur in the base metal precipitate settler where flocculent is added to speed up the settling process. The solids are removed in the thickener while a portion of the underflow (typically 80%) is recirculated back to the untreated feed solution to facilitate growth of copper sulphide crystals during the precipitation step to enhance the precipitation rate and settling. The excess metal sulphide solids in the underflow are neutralized, washed, filtered, dried and sold to a smelter as a valuable by-product (Botz *et al*., 2011).

The acidic thickener overflow (copper free cyanide solution) is sent to the neutralization reactor where the pH is raised to between 10 and 11. This is done by the addition of lime ($Ca(OH)_2$) and is necessary to convert aqueous hydrogen cyanide to the non-volatile cyanide anion represented by the following reaction (Barter *et al*., 2000; Botz *et al*., 2011; Ford *et al*., 2008):

$$2HCN_{(aq)} + Ca(OH)_2 \rightarrow Ca(CN)_2 + 2H_2O$$

As a result of the addition of sulphide during acidification and calcium during HCN neutralization by lime a gypsum type precipitate is formed. This is represented by the reaction below (Barter *et al*., 2000; Botz *et al*., 2011; Ford *et al*., 2008):

$$SO_4^{2-} + Ca(OH)_2 \rightarrow CaSO_{4_{(s)}} + 2OH^-$$

It is required that this precipitate is removed from the neutralized cyanide solution prior to the recovery of the precious metals in the downstream processes, as excessive scaling and fouling of the carbon can occur. This is done by adding a flocculant to the solution and sending it to a gypsum thickener where again a portion of the solid underflow is recycled to enhance the precipitation rate and settling. The gypsum can then be filtered dried and sold as a by-product (Botz *et al*., 2011; Ford *et al*., 2008).

As a major part of the SART process is operated at a pH below 5, the cyanide is present as hydrogen cyanide (HCN), which, if volatized, is in the most toxic form of cyanide. Mathematical models have been developed to predict the generation of HCN gas (Estay *et al*., 2012). This is a major concern with the SART process as the generation of HCN gas occurs at the reactors and thickeners/settlers. All four of these process units should be covered and sealed with vents to a lime scrubber system to capture the evolved HCN or H_2S gas and neutralize the emissions (Botz *et al*., 2011; Ford *et al*., 2008; Estay *et al*., 2012).

The final solution from the SART process contains low levels of base metals with free cyanide that has been liberated from the base metal cyanide complexes at a pH of about 10 and can be partially recycled back to the leaching stage in order to increase leaching performance or sent directly to the precious metal recovery processes downstream (Botz *et al*., 2011).

The leach solutions obtained from heap leaching of the Platreef ore will contain high copper and nickel concentrations, >60 mg/L (Mwase & Peterson, 2012), and the SART process can be used to reduce these concentrations to more acceptable levels, while also regenerating the associated cyanide for re-use in the leaching stage. This would hold multiple benefits for the downstream processing options and process economics.

2.5.3 Merrill Crowe/Zinc Cementation

A widely used option for the recovery of gold and silver from cyanide solutions is the well-known Merrill Crowe zinc cementation process (Hampton, 2002; Walton, 2005; Marsden & House, 2006). A review of this process is given below.

2.5.3.1 Overview

The classical method for the recovery of gold and silver from dilute cyanide leach slurries or clarified solutions is the process of cementation using zinc, either in the form of dust/powder or shavings, and was developed in the late 1880's and early 1890's. This process has later become known as the Merrill Crowe process (Hampton, 2002; Walton, 2005; Marsden & House, 2006).

This process was used almost exclusively for the direct recovery of gold and silver from cyanide leached solutions generated from either a solid-liquid separation step upstream, originating from a grinding and leaching operation, or from heap-leaching operations until the late 1970's, when carbon adsorption processes for the recovery of gold from leached solutions was widely adopted (Hampton, 2002; Walton, 2005).

Since the adoption of carbon adsorption processes the Merrill Crowe process and electrowinning were mainly applied to the more concentrated eluent generated from such a carbon adsorption or ion exchange resin process to recover the value material, but still amounts to about 25 - 30% of the worldwide gold production (Hampton, 2002; Marsden & House, 2006).

The Merrill Crowe process is, however, still preferred when the gold/silver content of the ore is high, the mercury content of the ore is high, high concentrations of flotation reagents or other organics that affect carbon adsorption are present or when the ore bodies exploited requires a small scale operation (Walton, 2005; Marsden & House, 2006).

The Merrill Crowe process was initially performed in sloped boxes filled with zinc shavings. The cyanide solution containing the precious metals was first filtered through sand filters and then allowed to pass through the zinc shavings. Several problems were encountered using this method. Zinc shavings became coated that essentially inhibited further precipitation while zinc consumption was also found to be very high.

Three main improvements were then made to the process (Walton, 2005):

- Addition of lead salt that increased the activity of the zinc and reducing the tendency to form passivating layers to enhance precipitation
- Substitution of zinc shavings with zinc powder increased zinc dissolution and precipitation rates (C.W. Merrill)
- The introduction of a de-aeration tower to lower dissolved oxygen levels that if to high allows for re-dissolution of gold (T.B. Crowe)

Alongside these improvements the Merrill Crowe process has evolved over the years to a highly efficient process with gold recoveries from solution in excess of 98% (Marsden & House, 2006).

2.5.3.2 Process Description

A typical Merrill Crowe process consists of 6 main process steps or units, which follow on the cyanide leaching stage. Each step forms an integral part in the overall performance and efficiency of the process and is listed below (Hampton, 2002):

- Solid liquid separation – Counter-current decantation is normally used
- Pregnant solution clarification – Clarified to about 1 mg/L suspended solids
- De-aeration of the solution – Performed in packed towers (Crowe) under vacuum
- Zinc and Lead addition – Added using a mixing cone
- Precious metal precipitation – Precipitate forms in pipeline between pumps and filters
- Filtration – Precipitated metals are filtered using filter presses

It should be noted that the Merrill Crowe process can be applied to metal bearing cyanide solutions directly after leaching, if the concentrations are high enough, or to the concentrated solutions generated by carbon or resin elution. It can thus be used as a recovery process on its own or as an intermediate stage during precious metal recovery (Marsden & House, 2006).

In a typical Merrill Crowe process, schematically shown in Figure 2.2, the pregnant leach solution is received either from a solution pond of a heap leaching operation or an unclarified solution tank of a solid-liquid separation operation. This solution is clarified to around 1 mg/L suspended solids to ensure effective precipitation of gold and silver. The clarification is typically performed in two stages. The first stage is a rake clarifier that reduces the solution to about 25 mg/L suspended solids. The second stage, a pressure leaf-filter, reduces the solution clarity to the required 1 mg/L suspended solids (Hampton, 2002; Walton, 2005).

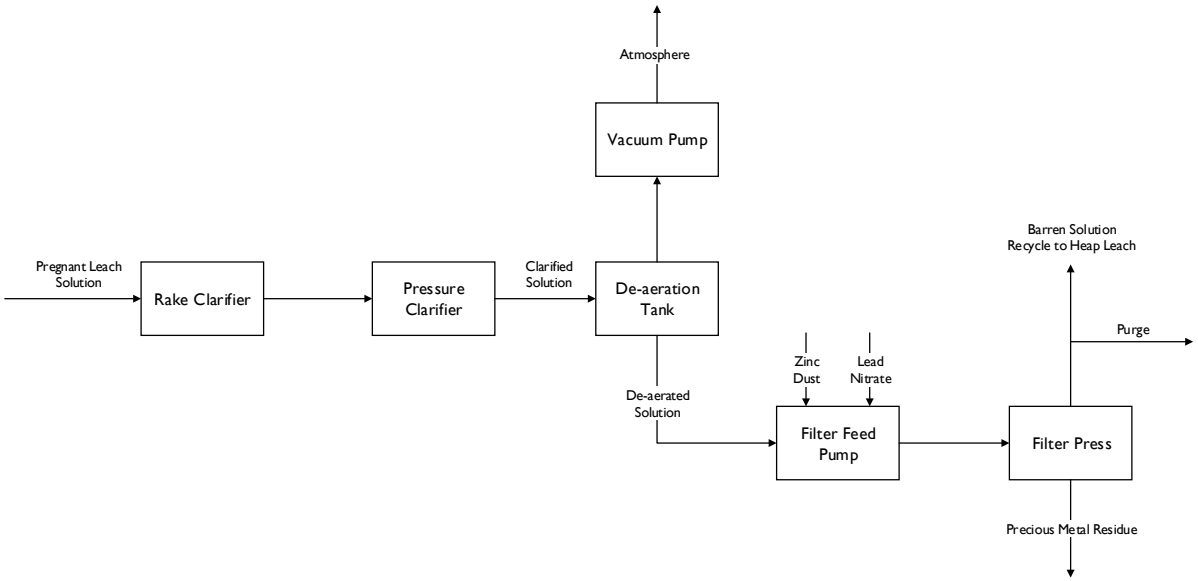

***Figure 2.2**: Merrill Crowe Block Flow Diagram*

The clarified solution is then transferred to the de-aeration tower (Crowe tower). A vacuum is applied at the top of the tower to reduce the dissolved oxygen content of the solution to less than 1 mg/L (Walton, 2005). The presence of dissolved oxygen reduces the precipitation kinetics as the reaction of oxygen reduction competes with that of gold. The rate of independent zinc dissolution also increases with increasing dissolved oxygen levels allowing for the potential of gold re-dissolution. It is thus a very important step in the process although not required when treating hot carbon strip solutions because of the low solubility of oxygen at the normal operating conditions of carbon elution (Hampton, 2002; Marsden & House, 2006).

The clarified pregnant cyanide solution enters the tower at the top and is distributed over a bed of packing to allow the formation of a thin film to release dissolved oxygen. The oxygen is removed through the vacuum pump while the de-aerated solution is discharged at the bottom (Hampton, 2002).

From here the solution flows out to the filter press feed pump. The zinc is added through a barren cyanide solution containing zinc powder, entering the piping ahead of the pump suction (Walton, 2005). The amount of zinc required for effective precipitation increases with decreasing precious metal concentration as the rate limiting step in the precipitation process is the diffusion of the metal cyanide complexes to the zinc surface (Hampton, 2002). The zinc is responsible for the reduction of the precious metals to their metallic state (precipitation). The reduction of gold, platinum and palladium are given by the following reactions (Walton, 2005; Marsden & House, 2006; Mpinga *et al.*, 2013):

$$2Au(CN)_2^- + Zn \rightarrow 2Au + Zn(CN)_4^{2-}$$

$$Pt(CN)_4^{2-} + Zn \rightarrow Pt + Zn(CN)_4^{2-}$$

$$Pd(CN)_4^{2-} + Zn \rightarrow Pd + Zn(CN)_4^{2-}$$

If free cyanide is available the direct transfer of cyanide ions, in the case of gold precipitation, is not required (Marsden & House, 2006):

$$Au(CN)_2^- + Zn + 4CN^- \rightarrow Au + Zn(CN)_4^{2-} + 2CN^-$$

There are several competing reactions that can result in poor precipitation and uneconomical performance, such as the potential of zinc to form zinc hydroxide: (Hampton, 2002):

$$2Zn + 8CN^- + O_2 + 2H_2O \rightarrow 2Zn(CN)_4^{2-} + 4(OH)^-$$

$$Zn + 4CN^- + 2H_2O \rightarrow 2Zn(CN)_4^{2-} + 2(OH)^- + H_2$$

$$2Au(CN)_2^- + Zn + 3(OH)^- \rightarrow 2Au + HZnO_2^- + 4CN^- + H_2O$$

$$Zn + 2OH^- \rightarrow Zn(OH)_2$$

This insoluble zinc hydroxide forms a passivating layer around the zinc inhibiting precious metal precipitation. It further increases the zinc consumption and greatly affects the overall efficiency of the process. It is thus important to maintain the correct level of dissolved oxygen and to optimize the level of free cyanide in solution as well as the amount of zinc dust added as all these factors play an integral part in limiting these competing reactions (Hampton, 2002).

The addition of lead nitrate (at the same stage as zinc) has been found to have a beneficial effect on the overall efficiency of the cementation process (Hampton, 2002). The lead is reduced at the zinc surface forming cathodic areas on which the precious metals are preferentially precipitated, resulting in more localized metal deposition that prevents the entire surface of the zinc to be coated with precious metals which allows for the dissolution of the zinc. Lead thus increase the activity of the zinc and further reduces the potential for the zinc to form passivating layers of zinc hydroxide. Excess amounts of lead, however, result in complete coating of the zinc thus preventing cementation (Hampton, 2002; Marsden & House, 2006).

The precipitation of the precious metals occurs within the pipeline after zinc addition and the suspended solid solution is fed to the filter press where the precipitated precious metals, as well as excess zinc and other precipitated base metals, are removed periodically. It is believed that much of the precipitation occurs during the filtration step as the solution passes through the filter cake. The precipitate may then be retorted, dried or calcined ahead of refining, depending on its mercury and base-metal content (Hampton, 2002).

The barren solution from the filter press can be recycled to the heap leaching stage or to the carbon elution stage to reduce cyanide consumption. A portion of this stream should, however, be bled out to ensure that zinc does not build-up to unwanted levels within the process (Hampton, 2002; Marsden & House, 2006).

For the process under development Merrill Crowe precipitation could be used, either directly after heap leaching or applied to carbon or resin eluates. A major benefit of heap leaching the Platreef ore is that the costly solid-liquid separation step is not required. The precious metal concentrations within this leach solution are, however, low which could lead to excessive reagent usage that negatively impacts the process economics. This could, however, be overcome if applied to the more concentrated carbon or resin eluates.

Current research (Desmond *et al*., 1991; Mpinga *et al*., 2013) has, however, shown that the recoveries of platinum and palladium from concentrated solutions by zinc cementation is poor and it has been suggested that other methods, such as carbon or resin adsorption and elution followed by electrowinning or thermal decomposition of the cyanide complexes, be explored.

2.5.4 Adsorption onto Activated Carbon

The hydrometallurgical process options for the treatment of gold and silver ores remained unchanged for the first 70 years of the twentieth century as gold plants used cyanide leaching followed by solid-liquid separation with the pregnant leach solution being treated by zinc cementation to precipitated the gold and precious metals from solution (Fleming, 2002). This process had its complications and research was conducted on developing a new process for the recovery of gold form cyanide leach solutions, i.e. adsorption onto activated carbon.

2.5.4.1 Overview

The ability of carbon to adsorb gold from cyanide leach solutions has been known since the middle of the 19th century. Johnson (1894) patented the use of wood charcoal for the recovery of gold from cyanide solutions (Fleming, 1992). It was, however, not used commercially until 1973 when the Homestake's Lead Gold Mine in South Dakota installed a carbon-in-pulp (CIP) plant for the recovery of gold from leach solutions (Laxen *et al*., 1979; Marsden & Fuerstenau, 1993).

Possible factors for the retarded development of the carbon adsorption process were the fact that there was no procedure for the elution of the gold from the carbon or the regeneration of the recycled carbon and that the Merrill Crowe process was believed to be adequate (Laxen *et al*., 1979).

Mintek and Anglo American Research Laboratory (AARL) in South Africa did, however, in the early 1970's preform significant research on the successful elution of gold from activated carbon, which opened the way for the commercialization of the CIP process (Davidson & Duncanson, 1977). The process at the Homestake Mine was very successful and changed the image of the CIP process and resulted in the widespread adoption of this new technology. CIP is now the process preferred for the recovery of gold from leach solutions and accounts for over 60% of the total world-wide gold production (Marsden & House, 2006).

The reasons for this remarkable turnaround in process selection in such a short period of time can be attributed to the fact that the economic analyses that was done proved that carbon adsorption was more economically favourable while the carbon processes further showed to be more robust and extremely versatile, both mechanically and chemically. Carbon adsorption processes were applied to almost any feed, still achieving high gold recoveries at lower capital and operating costs as opposed to the Merrill Crowe Zinc precipitation process (Fleming, 2002).

The fundamental difference between carbon adsorption and zinc cementation is that gold reduction on zinc dust requires a clarified solution whereas CIP processes can extract gold directly from pulp or slurries. This proves to be a major point of interest as the costly processes of filtration and solid-liquid separation are replaced by cost effective CIP plants. It was estimated that savings of between 20 - 50% could be expected (Fleming, 1992).

Carbon adsorption is, however, not without its weaknesses. The affinity of activated carbon is much higher for gold than for silver. Ores treated for its silver content would be better suited to the conventional Merrill Crowe process. The wide adoption of the carbon adsorption processes has led to the fact that some ore bodies were treated with potentially not the best flowsheet option (Fleming, 2002).

Activated carbon is further susceptible to chemical poison, fouling caused by flotation reagents or other chemicals within the solution. It is also known to adsorb base metal complexes, such as copper cyanide, in favour of gold, reducing the performance of the overall plant and diluting the product grade. The elution and regeneration stages of the carbon are further relatively high in cost (Stange, 1999).

2.5.4.2 Process Description

There are three basic carbon adsorption processes used for the recovery of gold from cyanide leach solutions (Fleming, 2002). These processes are:

- ***Carbon-in-pulp (CIP):*** Crushed ore is mixed with a cyanide solution in agitated tanks to form a slurry (pulp). After several leaching tanks the slurry is counter-currently contacted with activated carbon in several adsorption tanks. The carbon is then removed by screening once loaded with the gold.

- ***Carbon-in-leach (CIL):*** Similar to the CIP process. The only difference is that the carbon is mixed with the leachate solution and not the slurry, resulting in much less abrasion and ultimately carbon loss.
- ***Carbon-in-columns / Carbon-in-solution (CIC / CIS):*** In this process the cyanide solution is collected after a heap leach stage and passed through columns filled with activated carbon. The carbon is moved counter-currently as to promote the adsorption of the gold. This process is highly effective with regards to low grade ores.

In the process under development the Platreef ore body, which is very low in PGM and gold grades, is to be treated via a cyanide heap leaching process and if the resulting solution is to be processed in a carbon adsorption circuit, a process analogous to the gold carbon-in-solution process will be followed. This process is discussed further.

In a typical carbon-in-solution process, schematically shown in Figure 2.3, the pregnant leach solution is pumped upwards through several columns filled with activated carbon. The solution is pumped at a flowrate that is sufficient to lightly fluidize the bed of activated carbon (Fleming, 2002). The gold aurocyanide complex is readily absorbed onto the activated carbon. It is essential that high recoveries are achieved, as the solution leaving these columns will be lost to the tailings. It is thus required that the concentration of gold in solution leaving the adsorption circuit is between 0.001 and 0.02 mg/L responding to recoveries between 90 - 99% depending on the feed solution tenor, leading to acceptable economics (Stange, 1999).

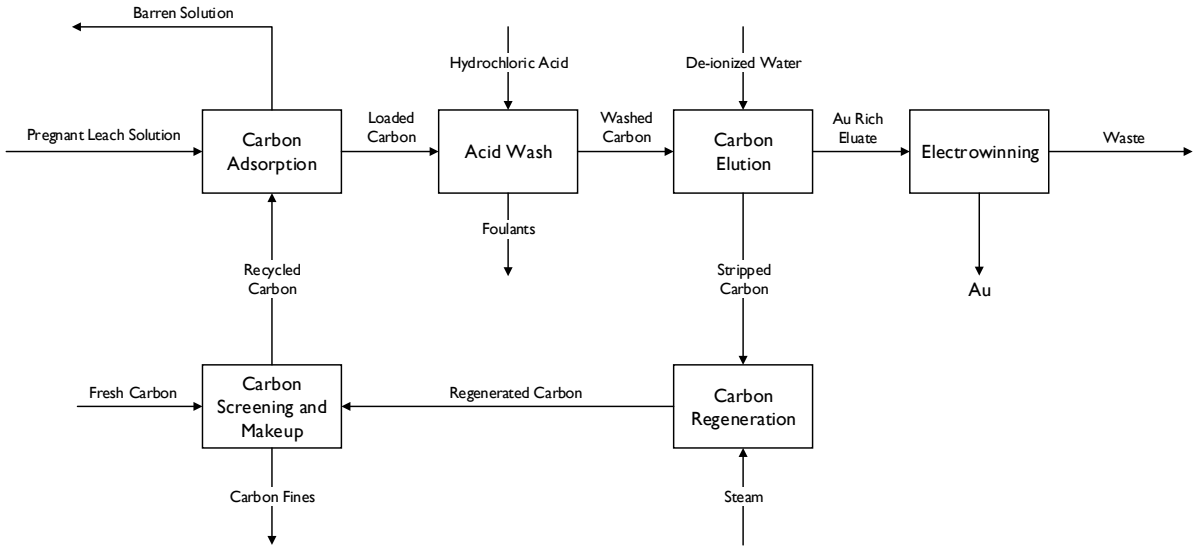

Figure 2.3: *Carbon-in-Solution Block Flow Diagram*

The flow of solution and carbon is counter-current and the gold concentration in solution thus decreases as it moves through the circuit, while the gold adsorbed onto the activated carbon increases as the carbon moves up the columns.

This setup is responsible for high recoveries as the low grade solution entering the last adsorption stage is in contact with fresh carbon. The loaded carbon is then screened as it leaves the adsorption stage, removing any fines generated during the adsorption cycle. Typical gold loadings of about 300 - 10 000 g/t are achieved which corresponds to a concentrating factor or upgrading ratio of between 600 - 1 500 (Fleming, 1992; Stange, 1999).

The barren cyanide solution leaving the adsorption stage is either disposed of or recycled back to the heap leaching stage. A portion of such a recycle stream should, however, be bled to ensure that iron, copper and other inert species not fully adsorbed onto the carbon do not build up to unwanted levels within the process.

Poor gold adsorption may be encountered due to the passivation of the carbon surface with calcium carbonate. The calcium comes from the addition of lime to keep the pH above 10.5 to ensure that a minimum amount of cyanide is lost as the toxic gas hydrogen cyanide (HCN), while the carbonate comes from the oxidation of cyanide. The loaded carbon can be acid washed to remove the calcium carbonate that precipitated onto the carbon and other inorganic foulants, while also cleaning fines out of carbon pores (Davidson & Schmidt, 1986; Fleming, 1992; Fisher & LaBrooy, 1997; Stange, 1999).

This acid wash step is performed by washing the carbon with a hot or cold solution of hydrochloric acid and increases elution efficiencies as base metal complexes are also removed (Davidson & Schmidt, 1986; Fleming, 1992; Fisher & LaBrooy, 1997). Acid washing can either be performed before or after the elution and regeneration stages. Acid washing prior to these two processes holds in various benefits for the overall performance of the plant. These benefits are listed below (Davidson & Schmidt, 1986):

- Elution efficiencies under acidic conditions and the removal of the passivating layer of calcium, blocking carbon pores, greatly increases the ease of gold elution
- Acid washing also strips excessive amounts of nickel, zinc and iron from the carbon, reducing reagent requirements to elute these metals effectively

- Acid washed carbon can be eluted to the same extent as non-acid washed carbon with the use of lower quality water
- Calcium, iron and magnesium are well known oxidants, which if loaded onto the carbon fed to the regeneration circuit could result in significant carbon losses
- The removal of silica will prevent the formation of refractory crystalline or sintered silicates on the carbon surface during thermal regeneration. These silicates are responsible for carbon fouling which negatively impacts the gold adsorption

Boshoff (1994) has, however, reported that the hot acid washing of carbon loaded with nickel can be detrimental to the performance of the elution and adsorption sections. The hot acid wash is responsible for stripping the $Ni(CN)_4^{2-}$ complex resulting in the formation of the $Ni(CN)_2$ precipitate (Jones *et al*., 1998; Boshoff, 1994).

This precipitate blocks the carbon pores and adversely affects the gold elution efficiencies. Boshoff (1994) further stated that this problem could be overcome by the introduction of cyanide in the elution step. The cyanide re-dissolves the nickel precipitate and allows for the effective elution of gold. To overcome this problem cyanide pre-treatment of the carbon has become normal practise in Anglo American Research Laboratory (AARL) elution, discussed below.

The next step in the process is the elution of gold from the activated carbon. During this stage the gold is desorbed from the carbon and dissolution of the gold occurs in the strip solution (Fleming, 1992). This can either be accomplished by two well established methods, the Zadra elution procedure or the Anglo American Research Laboratory (AARL) elution procedure (Ruane, 1982; Fleming, 1992; Marsden & House, 2006).

In the Zadra procedure, the carbon is eluted with a hot 95 - 100°C aqueous solution, containing 1% sodium hydroxide and 0.2% sodium cyanide, at atmospheric pressures. The solution is recycled through the loaded carbon to a subsequent electrowinning stage where the gold is plated onto steel wool cathodes. The solution leaving the electrowinning stage is recycled through until the eluate gold concentration is below 2 - 3 mg/L. This process usually takes up to 48 hours to complete, but can be reduced to about 12 hours by operating at higher temperatures (135 - 140°C) and higher pressures (500 - 600 kPa) (Ruane, 1986).

The Zadra elution method has a few disadvantages, listed below (Ruane, 1986):

- The required elution cycle times are relatively long
- Build-up of contaminants can severely restrict the elution efficiencies
- Water balance can be a problem due to electrowinning performed at close to 100°C

The second elution method is the AARL procedure in which the loaded carbon is first pre-treated for about an hour with 0.5 - 1 bed volumes of a 1% sodium hydroxide, 5% sodium cyanide solution followed by the elution with pure deionized water at a flow rate of about 2 bed volumes per hour. This method is based on the fact that an increase in ionic strength increases adsorption, thus deionized water would enhance desorption (Laxen *et al.*, 1979; Fleming, 1992). Elution is performed at temperatures above 110°C and pressures of 200 - 300 kPa, and takes 8 - 10 hours to complete (Runae, 1986; Marsden & House, 2006). The gold is then recovered via electrowinning while the barren solution is sent to the tailings dam for disposal (Fleming, 1992; Stange, 1999; Marsden & House, 2006).

The stripped carbon is then first screened, to remove any undersized carbon from the process, where-after it is regenerated in a rotary kiln, at temperatures in the range of 650 - 750°C, to either carbonize or volatilize any organic compounds that could have adsorbed onto the activated carbon. This step is carried out in the presence of steam to minimize carbon degradation due to oxidation (Fleming, 1992; Stange, 1999). Thermal regeneration of the carbon restores it to its original activity for re-use in the adsorption stage, as is required to ensure that excessive carbon fouling does not occur.

The regenerated carbon is then fed back to the adsorption stage where the whole process is repeated. Fresh carbon is added to make up for any carbon loss, resulting from the abrasion of the carbon particles during physical transport and the fines produced as a result of the thermal and osmotic shock during regeneration (Zanitsch & Lynch, 1978).

2.5.4.3 AAC Pump Cells vs. Conventional CIP tanks

A relatively new development in carbon adsorption, and probably the most significant in recent years, has been the research conducted by the Anglo American Corporation of South Africa, which during the late 1980's developed an alternative approach to the design of a CIP circuit (Dippenaar & Proudfoot, 2005).

This new development, called AAC Pump Cells, was based on the fact that improved metallurgical performance can be achieved by utilizing the carousel mode of operation, which involves the rotation of the solution or pulp feed to the adsorption contactors filled with discrete amounts of carbon. The carbon is managed in a batch format, eliminating interstage transfer and inventory balancing (Schoeman *et al*., 1996; Rogans & McArthur, 2002.

The counter-current movement of solution relative to carbon is thus simulated by altering the solution feed and discharge positions, without the need of pumping the carbon from one stage to the other (Whyte *et al*., 1990; Schoeman *et al*., 1996; Rogans & McArthur, 2002).

There are several benefits of using the carousel method as opposed to the normal cascade mode of operations. No backmixing of carbon, which results in the dilution of the gold solution tenor, occurs in the carousel mode of operation. The rate of gold adsorption is directly influenced by the gold concentration in solution, and thus higher gold adsorption rates can be achieved that leads to higher gold on carbon loadings (Rogans & McArthur, 2002).

The AAC Pump Cell mode of operation further operates at higher carbon concentrations per stage. As the rate of gold adsorption is a function of the total surface area of the carbon (Rogans & McArthur, 2002), lower solution residence times, and higher carbon concentrations are used. This is believed to increase the rate of gold adsorption. AAC Pump Cells operate at solution residence times of between 15 - 20 minutes at carbon concentration of between 20 - 60 g/L (Rogans & McArthur, 2002; Dippenaar & Proudfoot, 2005; Fleming *et al*., 2011), while conventional CIP circuits operate at solution residence times of about an hour at carbon concentrations of below 25 g/L (Stange, 1999; Fleming *et al*., 2011).

Mathematical modelling and plant experience has shown that the amount of carbon per stage determines the gold extraction efficiency (Buson *et al*., 1999) and not the solution or pulp residence time. Fleming *et al*. (2011) has shown that when the size of the adsorption tanks is halved, the same gold in solution profiles are achieved. Solution or pulp residence time thus has no significant influence on the metallurgical performance of a CIP circuit. This is the principle that is used in the design of the AAC Pump Cells.

Significant savings can be achieved when installing a Pump Cell circuit as opposed to a conventional CIP circuit. Overall capital cost savings of up to 20% and operating cost savings of up to 50% can be achieved (Stange, 1999; Rogans & McArthur, 2002).

These savings are primarily due to the reduced size of the adsorption contactors and the benefits of moving less carbon through the plant (smaller elution and regeneration sections) as a result of the higher gold on carbon loadings that can be achieved. Loadings achieved at the AngloGold's West Wits mine were almost doubled after the installation of an AAC Pump Cell circuit (Rogans & McArthur, 2002).

Figure 2.4 is a comparison between the relevant performance of several plants that utilises either the conventional CIP or AAC Pump Cell mode of operation, based on their respective gold upgrade ratios, i.e. the carbon to solution flowrate. This data was collected from several studies (Whyte *et al.*, 1990; Rogans & McArthur, 2002; Dippenaar & Proudfoot, 2005; Fleming, 2002).

It can be seen that the upgrade ratios obtained in the AAC Pump Cell circuits are superior to that of the conventional CIP circuits, as a result of the benefits associated with the carousel mode of operation. The elimination of backmixing and higher mixing efficiencies due to the smaller contactors result in higher overall gold adsorption rates, maximising gold on carbon loadings.

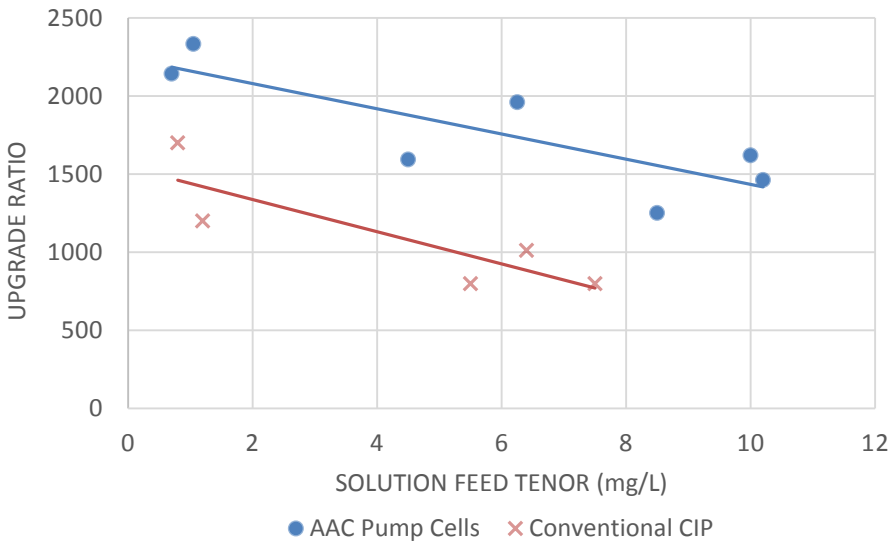

Figure 2.4: *Upgrade ratios achieved in AAC Pump Cell circuits vs. Conventional CIP circuits (Whyte et al., 1990; Rogans & McArthur, 2002; Dippenaar & Proudfoot, 2005; Fleming, 2002)*

A higher upgrade ratio has drastic implications on both capital and operating costs as less carbon is being transferred to the subsequent elution and regeneration circuits leading to smaller equipment needed to obtain the same outcome. Carbon loaded to higher loadings may require longer elution times but the operating costs associated with this is far more inferior to the savings obtained from operating with the use of the AAC Pump Cells.

Taking all of these benefits into consideration it is proposed that the AAC Pump Cell circuit design and mode of operation be utilized if adsorption onto activated carbon is used in processing the Platreef ore deposit.

2.5.5 Ion Exchange Resins

In recent years the developments of ion exchange resins for the extraction of gold from cyanide leach solutions has led to the fact that resins has started to replace carbon in some of the modern gold recovery plants (Kotze *et al*., 2005). Recovery of gold by resins offer additional advantages as opposed to carbon and is discussed in the next sections.

2.5.5.1 Overview

Since the development of cyanidation treatment of gold ores in the early 1880's very few advances had been made as to the technology that can be used for gold recovery till the development of the adsorption of activated carbon came along in the 1970's (Fleming, 1992; Fleming, 2002; Kotze *et al*., 2005).

Detailed studies as to the use of ion exchange resins were done in the early 1950's and the subsequent successful commercialization of the activated carbon process had given new interest into the development of resin technology. Although resins were not used in the western world during the initiation of the carbon processes, the USSR practised it widely in its uranium industry (Fleming, 1992)

Significant developments around the gold-selective ion-exchange resins and the commercialization thereof has made this technology available for the western world today and is used as an alternative to the widely used carbon adsorption processes, as it is believed to hold in various additional advantages (Fleming & Cromberge, 1984*a*; 1984*b*, 1984*c*; Kotze *et al*., 2005).

Milbourne *et al*. (2003) has proposed the use of ion exchange resins for the recovery of PGMs from cyanide leach solutions. These resins include:

- ***Chelating resins***: Highly selective but expensive. Functional groups hold the desired metals by mechanisms similar to those which bind ligands together.
- ***Strong base resins***: Adsorbs anions including anionic complexes. Not as selective as chelating resins but is less expensive.
- ***Reducing resins***: High loadings of precious and platinum group metals can be achieved as the ions are reduced to metals within the resin matrix. Not easily strippable, thus destruction needs to be economically feasible.

2.5.5.2 Process Description

The recovery of gold by ion exchange resins, shown in Figure 2.5, follows the same basic process route as that of adsorption onto activated carbon. The same general adsorption process is used but different elution procedures are followed to strip the gold from the resin.

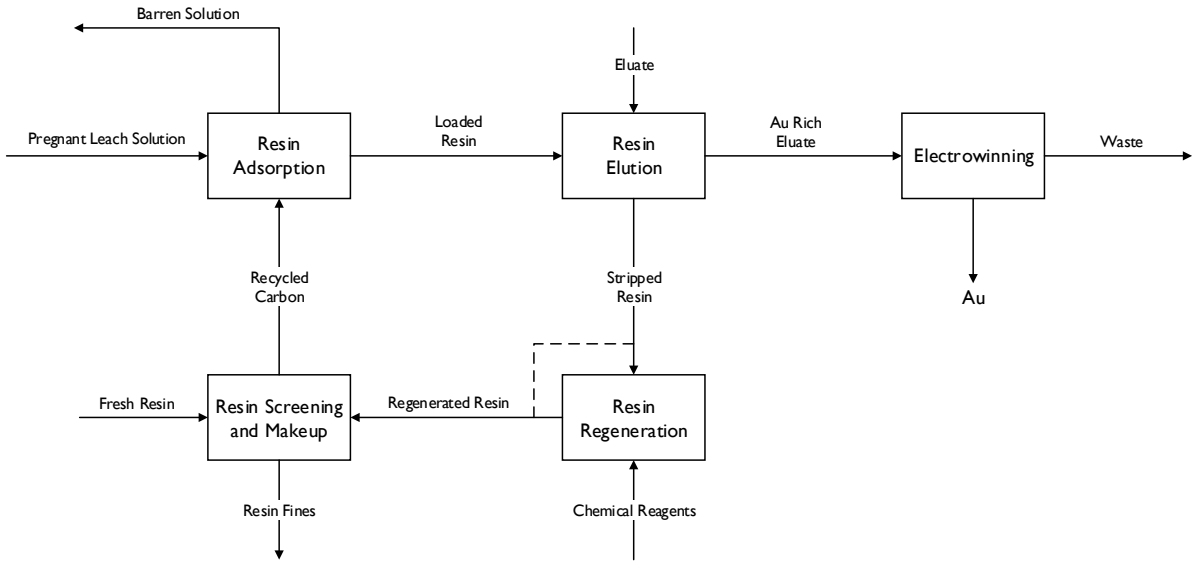

Figure 2.5*: Resin-in-Solution Block Flow Diagram*

There are mainly three different eluting agents used in eluting gold loaded onto ion exchange resins. These include zinc cyanide, thiocyanate and acidic thiourea (Fleming & Cromberge, 1984*a*, 1984*c*; Kotze *et al*., 2005; Marsden & House, 2006). These eluting agents use different mechanisms to strip the gold from the resin, either physical or chemical elution. Zinc cyanide and thiocyanate are physical eluting agents in that they physically displace the loaded aurocyanide ion from the resin by either the zinc cyanide or thiocyanate ion.

Acidic thiourea on the other hand is a chemical eluting agent in that it alters the aurocyanide ion in such a way that it is no longer adsorbed onto the resin. The advantages and disadvantages of these three different eluting agents are listed in Table 2.2 (Fleming & Cromberge, 1984*a*; Kotze *et al*., 2005; Marsden & House, 2006):

Table 2.2: *Advantages and Disadvantages of different resin eluting agents*

Zinc Cyanide	*Thiocyanate*	*Acid Thiourea*
Advantages		
• Effective in eluting all metal cyanide complexes	• Effective in eluting all metal cyanide complexes	• Very high gold elution efficiencies
• Resin can be regenerated to full capacity	• Resin can be regenerated to full capacity	• No regeneration of resin required
• Zinc cyanide can be recycled efficiently	• Nontoxic chemicals used for elution and regeneration	• Nontoxic chemicals used for elution
• Poses no problems regarding materials of construction	• Very fast elution kinetics	
Disadvantages		
• Resin regeneration required	• Resin regeneration required	• Slow elution kinetics
• Slow elution kinetics	• Corrosion of electrodes during electrowinning	• Corrosion of electrodes during electrowinning
• Toxic hydrocyanic gas produced during regeneration		• Decomposition of thiourea in acidic conditions
		• Ineffective base metal elution

Resins eluted with zinc cyanide requires a regeneration step as it is necessary to remove the loaded zinc cyanide from the resin prior to the recycling of the resin back to the adsorption section. The loaded zinc cyanide greatly hampers the ability of the resin to effectively adsorb the gold from solution. The zinc cyanide can, however, be stripped from the resin by means of acid regeneration, in which the resin is washed with a dilute mineral acid. The resulting solution can then be mixed with a caustic soda solution, regenerating the zinc cyanide for reuse in the elution cycle. This process has been shown to be effective in small scale and large industrial operations (Fleming & Cromberge, 1984*a*, 1984*c*; Kotze *et al*., 2005).

Thiocyanate has been shown to have improved elution kinetics over that of zinc cyanide, both for gold and base metals, but the fact that there is currently no establish method of recycling the thiocyanate loaded onto the resin has limited its application in industry (Fleming & Cromberge, 1984*a*, 1984*c*; Kotze *et al*., 2005). A process by which the thiocyanate can be recycled by washing the resin with a ferric iron solution has been shown to be effective but has not been implemented on large scale operations (Marsden & House, 2006).

Although acidic thiourea is the most effective eluting agent with respects to gold, the decomposition thereof in acidic conditions and the high cost thereof proves to be problematic from an economic standpoint. It is furthermore ineffective in eluting excessive amounts of base metals loaded onto the resin. Both these factors have shifted the balance towards zinc cyanide as the primary eluting agent (Fleming & Cromberge, 1984*c*).

It has thus been proposed, based on the above mentioned characteristics of each eluting agent, that acidic thiourea be used for the elution of gold from gold-selective resins, as the base metal elution efficiencies of this agent is poor, while zinc cyanide is proposed for the elution of gold loaded onto non-selective resins (Kotze *et al*., 2005).

The differences and possible additional advantages or disadvantages that resins may have over that of activated carbon are listed below (Fleming 1993; Fleming & Cromberge, 1984*b*; Kotze *et al*., 2005):

- Resin equilibrium loadings of aurocyanide are superior to that of activated carbon. A smaller inventory is thus required resulting in a smaller elution stage reducing operating and capital cost.
- Resins do not require the need for an acid wash stage as it is not susceptible to organic fouling, such as the passivating layer formed by the adsorption of calcium carbonate.
- The elution of activated carbon is normally performed using high temperatures (120 - 130°C) and increased pressures (200 - 500 kPa). Resins could be eluted at ambient pressures and at temperatures in the range of 50 - 60°C. Zinc cyanide, thiocyanate or acidic thiourea is typically used as oppose to sodium cyanide, sodium hydroxide and de-ionised water used in carbon elution.
- Activated carbon requires a thermal regeneration stage in which the carbon is “cleansed” prior to being recycled back to the adsorption stage. This step is unnecessary

when resins are used and could greatly affect the operating and capital costs based on the scale of the operation.

- Carbon is susceptible to chemical poison as flotation reagents, oils, solvents and lubricants inhibiting the loading of gold. Adsorption if further depressed by hematite and shales or clay-type minerals, while these have little effect on resins.
- Resins can further be tailor made for specific operations or applications, such as the gold selective Minix resin (Kotze *et al*., 2005). Different metals could thus be extracted using different resins at various stages within the process. It can be used to co-extract base metal cyanide complexes opening the possibility of cyanide recycling to the leaching stage, greatly reducing the operating costs of the process. Current available resins are on the other hand less selective for the aurocyanide ion over that of the base metal cyanide complexes.
- Resin particles are furthermore much smaller than that of carbon which could provide additional challenges with regards to interstage screening.

It can thus be seen that there are a few differences in the process flowsheets between a typical activated carbon (CIS) process and that of ion exchange resins (RIS). Each poses their own challenges which need to be overcome to ensure effective extraction and recovery in an economically favourable process.

2.6 Platreef Process Options Summary

From the preceding discussions several process flowsheet considerations and options were identified that could be used in processing the Platreef ore deposit which include cyanide heap leaching, base metal recovery and cyanide regeneration via the SART process and PGM recovery with the use of zinc precipitation via the Merrill Crowe process, adsorption onto activated carbon and adsorption via ion exchange resins. Several of these options, with specific application to the Platreef ore deposit, are discussed in further detail in the next chapter.

Chapter 3

Application of Gold Ore Processing Techniques to a Low Grade PGM Ore

This chapter focuses on the results obtained from concurrent experimental studies based on the application of several of the treatment options typically used in the gold industry, as discussed in Chapter 2, on a low grade PGM ore. Firstly the extraction of platinum group metals by heap leaching is discussed, followed by a discussion of the downstream recovery processes for recovering the PGMs from the pregnant cyanide leach solution.

3.1 Proposition

This project formed part of a major collaboration between several researchers at Stellenbosch University as well as the University of Cape Town. In the broader collaborative project, investigations were conducted to determine if well-established gold processing techniques could be applied or adapted for the extraction and recovery of platinum group metals from low grade ores from a heap leaching perspective. Each study formed an independent project and each specific technique or technology was evaluated individually for its use in the PGM industry. These projects, with the implicit proposition evident from the title, included:

- Evaluation of bio-heap leaching of base metals and cyanide heap leaching of PGMs and Au from a low grade PGM concentrate or crushed ore (Mwase, 2009)
- Evaluation of the extraction of base metals from the bio-leach solution by ion exchange resins and the subsequent elution thereof (Liebenberg, 2013)
- Evaluation of the extraction of PGMs from the cyanide leach solution by activated carbon (Mpinga, 2012)
- Evaluation of the elution of PGMs absorbed onto activated carbon from the cyanide leach solution (Snyders, 2011)

- Evaluation of the extraction of PGMs from the cyanide leach solution by ion exchange resins (Schoeman, 2012)
- Evaluation of the elution of PGMs absorbed onto ion exchange resins from the cyanide leach solution (Schoeman, 2012)
- Evaluation of the potential of zinc precipitation to recover the PGMs from the cyanide leach solution or carbon adsorption eluates (Mpinga *et al*., 2013)

The results obtained from these individual experimental studies are reviewed in the following sections.

3.2 Heap bio- and cyanide leaching of low grade PGM concentrate

Recovery of gold is most commonly achieved today by some means of cyanidation, in which the gold bearing ore is either heap leached or subjected to pressure leaching or tank leaching in carbon adsorption circuits. It is a well understood and well established technique and is almost exclusively used in industry for the recovery of gold from gold bearing ores (Fleming, 1992; Marsden & House, 2006).

As a result of this, interest has been shown in the recovery of platinum group metals from low grade concentrates using cyanide leaching as the primary extraction process (Mwase, 2009). The recovery of these metals from low grade ores and concentrates via cyanidation was mostly unknown up to the start of this project, as interest has only been shown in recovering these metals from recycled automobile catalysts using pressure cyanide leaching (Desmond *et al*., 1991; Kuczynski *et al*., 1992; Roijals *et al*., 1996).

Mwase (2009) performed extensive work on evaluating the application of cyanidation on recovering platinum group metals from low grade Platreef flotation concentrates. As the Platreef ore body mainly consists of base metals, a sequential leaching technique was proposed, in which the base metals are first extracted with the use of a bio-leach followed by the extraction of the PGMs in a subsequent cyanide leach of the bio-leach residue (Eksteen *et al*., 2013; Mwase *et al*., 2012*a*, 2012*b*).

Mwase *et al*. (2012*b*) have shown that bio-leaching of Platreef concentrate is effective in recovering most of the base metals, as recoveries of 91.1%, 98.5% and 83.5% for Cu, Ni and Co after 88 days of leaching at 65°C were achieved.

Iron recovery was low, at only 38.4%, and was attributed to the fact that the iron was mostly present as jarosite. Additional test work confirmed that conditions can be altered to minimize jarosite formation. It was further found that no significant dissolution of platinum, palladium and gold occurred during the bio-leach, however, rhodium and ruthenium leached to significant quantities. This implies that these two metals should be recovered from the sulphate solution.

The resulting bio-leach residue was then rinsed, lime washed and restacked for a subsequent cyanide leach to recover the precious metals (Mwase *et al*., 2012*b*). Two separate tests were conducted on two batches of bio-leach residue. The recoveries of the precious metals after 45 days of the cyanide leach are presented in Table 3.1.

Table 3.1: *Precious metal extraction by cyanide leaching after 45 days (After Mwase et al., 2012b)*

Test	*Recovery (%)*		
	Pt	***Pd***	***Au***
A	34.3	96.5	63.4
B	32.2	92.5	97.5

From Table 3.1 it can be seen that palladium recoveries were very high in both tests, while gold recoveries showed significant variations. It was believed that mineralogical factors may have led to these inconsistencies. Platinum recoveries were, however, low in comparison to that of palladium and gold. This was attributed to the fact that the platinum was mainly present in the concentrate in the form of sperrylite ($PtAs_2$), known to be refractory to direct cyanidation. The observed platinum recovery curves were found to be linear and it was expected that an increase in recovery could be achieved by increasing the leaching cycle.

Based on the promising results of the aforementioned studies, a subsequent study was performed on the sequential heap leaching of Platreef ore (Mwase *et al*., 2014). Two separate ore samples with different size distributions of -25mm + 1mm and -6mm +1mm were used in these experiments. The ore samples were first bio-leached at 65°C in columns, simulating a heap leach, followed by a cyanide leach at 50°C.

The best results were obtained for the smaller size fraction ore sample and after 304 days 93%, 75% and 53% of Cu, Ni and Co were extracted by bio-leaching. The subsequent cyanide leach extracted 57.8%, 99.7% and 90.3% of platinum, palladium and gold after 60 days. Palladium and gold were found to leach very fast, while platinum showed the same retarded leaching kinetics as observed with the leaching of concentrate, reported by Mwase *et al*. (2012*a*, 2012*b*).

From these results it was proposed that sequential heap leaching, bio-leach followed by cyanide leaching, of Platreef ore could be a potential alternative treatment route for the recovery of precious metals contained within this ore body.

It has also been proposed that a possible two stage cyanide leach be performed as a result of the retarded leaching kinetics of the platinum. A short term leach can be performed in which about 35% of the majority of the palladium and gold are extracted, followed by a long term leach to extract some of the remaining platinum.

Based on the results obtained from the studies conducted on the sequential heap leaching of low grade Platreef ore concentrate (Mwase, 2009; Mwase *et al*., 2012*a*, 2012*b*) several other studies was initiated to evaluate different techniques for the recovery of the precious metals from the resulting leach solutions (Mpinga, 2012; Mpinga *et al*., 2013; Schoeman, 2012; Snyders *et al*., 2013). These studies are reviewed in the next sections.

3.3 Adsorption of PGMs onto activated carbon

In an attempt to assess the use of activated carbon for the extraction of platinum group metals from cyanide leach solutions, a study by Mpinga (2012) was undertaken to determine whether activated carbon could accomplish this task, in analogy to the well-known gold CIP process.

Statically designed experiments were performed using synthetic alkaline cyanide solutions that closely represented those expected from a heap leach solution. The adsorption efficiencies of the PGMs were studied as a function of solution pH, base metal concentration, free cyanide concentration, thiocyanate concentration and initial precious metal concentration.

Mpinga (2012) found that platinum, palladium and gold can be effectively adsorbed onto activated carbon within 2 hours, at which recoveries in excess of 90% were found. Extraction of the precious metals proceeded after this point, but at much slower rates until pseudo equilibrium was reached.

Nickel was found to be extracted at the same rate as that of the precious metals (90% after 2 hours) while copper adsorption was much slower at less than 30% adsorbed after 2 hours. It was suggested that copper recovery can be minimised by shortening the adsorption residence time.

Optimum adsorption conditions were found to be at a pH of 9.5, with the lowest level of free cyanide and base metals present in the feed solution. Precious metal recoveries after two hours at these conditions were 98%, 92% and 100% for platinum, palladium and gold respectively, with the carbon showing the greatest affinity for gold, followed by platinum and palladium.

It was concluded that activated carbon has the potential to adsorb platinum and palladium from dilute cyanide solutions, and that further research is required, both fundamental and economic approaches, to assess its use in the PGM industry.

3.4 Elution of PGMs from activated carbon

When activated carbon was first proposed as a treatment option for the recovery of gold, the feasibility of stripping the gold from the carbon first had to be shown to be effective to allow for the reactivation and reuse of the carbon (Heinen *et al*., 1978; van der Merwe, 1991).

The same problem has been presented with regards to the application of carbon technology for the recovery of platinum group metals. The feasibility of eluting these metals from loaded activated carbon with the well-known AARL elution method was thus investigated (Snyders, 2011; Snyders *et al*., 2013; Eksteen *et al*., 2013).

Carbon loaded with various amounts of platinum, palladium and gold was eluted in these studies. The loaded carbon was pre-treated with a sodium cyanide sodium hydroxide solution. Elution experiments were conducted at 60°C and 80°C, and it was found that platinum and palladium could be effectively eluted within 6 - 8 bed volumes. Temperature was found to play a significant role in the rate of elution, similar to that observed in gold elution operations. Increasing the elution temperature from 60°C to 80°C resulted in significantly faster elution kinetics. About 95% of the platinum and palladium were eluted after 5 bed volumes at 80°C, while only about 75% were eluted at 60°C. As gold elution is typically performed at 100 - 120°C (Marsden & House, 2006), it is assumed that even higher elution efficiencies of PGMs could be achieved at these higher temperatures.

Gold elution efficiencies were, however, low at these temperatures, with recoveries in the range of 35% after 10 bed volumes. A minimum temperature of 95°C has been reported (Greenhalgh *et al*., 1991) that is necessary for the efficient elution of gold (95%) and that the recovery of gold rapidly decreases at lower elution temperatures (only 25% at 85°C).

The observed gold elution recoveries were thus deemed to be in accordance with literature values and the effects that platinum and palladium could have had on gold elution efficiencies were deemed insignificant, if there was any.

Eksteen *et al*. (2013) reported that an optimum condition of the pre-treatment step exists, that ensures that maximum precious metal recovery is achieved during elution. The optimum pre-treatment solution was reported to be a solution containing 2% sodium cyanide and 0.55% sodium hydroxide. It was also stated that the ionic strength of the eluate plays a significant role in ensuring efficient PGM elution. The same trend to that observed in gold elution was observed, i.e. a decrease in the ionic strength of the eluate resulted in higher elution efficiencies (Eksteen *et al*., 2013).

Based on these findings it was concluded that the elution of PGMs loaded onto activated carbon is viable and that it could pave the way for the application of carbon adsorption plants in the PGM industry.

3.5 Adsorption of PGMs onto ion exchange resins

In an attempt to assess the application of ion exchange technology to the recovery of platinum group metals from cyanide leach solutions, a study was undertaken to determine whether industrial available ion exchange resins could be used to accomplish this task, in analogy to the well-known gold carbon-in-pulp and resin-in-pulp processes (Schoeman, 2012).

Screening test were performed to assess the ability of different ion exchange resins, both weak and strong base, to extract platinum group metals from dilute cyanide leach solutions. Synthetic cyanide solutions, containing both base and precious metals were prepared. From the subsequent screening tests four different ion exchange resins were identified, all strong base resins, that showed particular preference to adsorb the precious metals from solution. These resins included Amberjet 4600, Dowex 21K XLT, Amberlite PWA 5 and the developmental Dowex Minx XZ 91419 resin and were selected based on their PGM recovery potential and to assess the difference that different exchange capacities and functional groups could have on PGM recovery.

Batch kinetic tests, isotherm evaluations and repeated loadings experiments were performed using these resins. It was found that the exchange capacity and functional groups of the resins played a significant role in their affinity for the different metals in solution. The Amberjet 4600 and Dowex 21 K XLT resins, both containing higher exchange capacities and smaller functional groups, extracted almost all metals in solution in a non-selective manner.

The resins with smaller exchange capacities and larger functional groups, Amberlite PWA 5 and Minix XZ 91419, showed higher affinities for the lower charged platinum, palladium and gold anions, over that of the higher charged base metal anions. As a result of the higher selectivity of these two resins for the precious metals, further test work was performed.

It was found that changes in the temperature and pH of the solutions did not have significant effects on the loadings of the precious metals onto Amberlite PWA 5 and Minx. An increase in temperature and initial concentration were found to enhance the adsorption kinetics of platinum, palladium and gold.

Selectivity series were developed with the use of repeated adsorption tests. It was found that Amberlite PWA 5 showed high selectivity towards the divalent anions of platinum ($Pt(CN)_4^{2-}$) and palladium ($Pd(CN)_4^{2-}$) over that of the monovalent aurocyanide ion ($Au(CN)_2^-$). Precious metal loadings were found to increase with each repeated contact, while nickel, copper and iron were rejected by the resin. Gold adsorption was, however, found to be very slow.

The Minix resin was shown to have the highest affinity for gold. All other metals were rejected by the resin, with no copper, nickel or iron loadings observed after 4 repeated contacts. Platinum and palladium loadings increased gradually up until a point where stationary loadings were observed. This resin thus provides a method for the selective adsorption of gold, even from solutions containing much higher base and platinum group metals.

From the work conducted by Schoeman (2012) it was concluded that ion exchange resins could be used in the simultaneous recovery of platinum, palladium and gold from dilute cyanide solutions, as recoveries in excess of 95% was observed. It was proposed that the Amberlite PWA 5 be used for the selective extraction of platinum, palladium and gold, while the more gold selective Minix resin be used for the extraction of gold.

3.6 Elution of PGMs from ion exchange resins

The ability of ion exchange resins to adsorb platinum, palladium and gold from dilute cyanide solutions prompted a subsequent study in the elution of these three metals loaded onto both Amberlite PWA 5 and Minix XZ 91419 (Schoeman, 2012).

Three different eluting agents, zinc cyanide, potassium thiocyanate and acidic thiourea, were investigated to assess whether these eluates, used successfully in the gold industry (Fleming & Cromberge, 1984*a*, 1984*c*; Kotze *et al*., 2005), can be used to strip the resins, loaded with high amounts of platinum, palladium and gold, to sufficiently low levels that would allow for the reuse of the resins in the adsorption circuits.

Schoeman (2012) found that effective elution of both platinum and palladium, loaded on to Amberlite PWA 5 and Minix, could be achieved with zinc cyanide and potassium thiocyanate, with recoveries in excess of 95% after 15 bed volumes of the eluates were used. Gold elution was, however, slow with only about 60% recovered in the same period.

Elution of these metals with acidic thiourea was also performed. It was found that gold and palladium, loaded onto both resins, could be stripped effectively at close to 99% recovery after 15 bed volumes. Elution of platinum with acidic thiourea was, however, found to be ineffective with no platinum being recovered even after 20 bed volumes of eluate was used. It was believed to be as a result of a kinetic reaction between the platinum cyanide ion and the thiourea, due to the low pH of the eluate solution.

Sensitivity analyses were also conducted on the effect of various parameters, such as temperature, eluate concentration and flowrate, on the elution efficiencies. It was found that the eluate concentrations had the biggest effect on overall elution performance. Optimum elution efficiencies were achieved with a 0.5 M zinc cyanide solution at 50°C and a flowrate of 5 BV/h, while optimum elution efficiencies with acidic thiourea were achieved with a 0.7 M thiourea 0.2 M hydrochloric acid solution at 60°C and a flowrate of 5 BV/h.

Based on the results obtained from this study, it was concluded that the elution of platinum group metals loaded onto ion exchange resins are viable, opening the way for further process development for the implementation of these technologies for use in the PGM industry.

3.7 Zinc precipitation of PGMs from cyanide leach solutions

Mpinga *et al*. (2013) conducted a study to evaluate the application of the Merrill Crowe zinc cementation process to a high grade cyanide solution, such as that produced during the elution of platinum group metals from activated carbon, as obtained in the work by Snyders *et al*. (2013). The process was evaluated for its use in the simultaneous recovery of platinum group metals from a solution containing high amounts of platinum, palladium and gold as well as the base metals copper and nickel.

Statistically designed experiments were performed in a mechanically stirred reaction vessel in the presence of pure nitrogen gas, by varying parameters such as the free cyanide concentration, temperature, reaction time and metal concentration.

Experiments were conducted on solutions containing 220 and 880 mg/L precious metals. Optimum recoveries were achieved at operating conditions of 60°C, 150 mg/L free cyanide and a reaction time of 90 minutes.

Extraction efficiencies were low with recoveries of 91%, 54% and 19% of gold, palladium and platinum respectively, even though 1.5 times the stoichiometric amount of zinc was added. When the zinc addition was increase to 3.4 times the stoichiometric requirement, platinum and palladium recoveries were only found to be 36% and 46% respectively.

Due to these low overall extraction efficiencies of platinum and palladium, it was proposed that an alternative method for the recovery of the precious metals from typical carbon adsorption eluates be investigated.

3.8 Recovery of PGMs from automobile catalysts

Independent studies conducted at the United States Bureau of Mines focused on the high temperature leaching of platinum group metals from automobile catalysts (Desmond *et al*., 1991; Kuczynski *et al*., 1992). One of the aims of these studies was to determine the most effective method for recovering the PGMs from solution. This was done by treating the resulting pregnant liquor with several of the aforementioned techniques to recover the PGMs from solution which included zinc and aluminium cementation, adsorption onto activated carbon, extraction via ion exchange and reduction resins, electrowinning and thermal decomposition of the cyanide complexes.

For the cementation tests leach solutions containing 550 mg/L platinum, 100 mg/L rhodium and 50 mg/L of free cyanide were treated by the addition of powdered zinc or aluminium. Additions of 1, 1, 2, 4 and 8 grams each were made at 1 hour intervals. Tests were repeated at 30°C, 45°C and 60°C and samples were taken after 1 hour and before each addition.

Tests conducted at 60°C using zinc yielded the best results with 50% removal of platinum and 10% removal of rhodium. Aluminium was found to be ineffective in the removal of both metals. Due to the fact that the PGM recoveries were so low additional tests were not conducted.

The loading of PGMs onto activated carbon was also investigated. Leach solutions were contacted with 10 times the recommended amount of carbon in a stirred beaker for 1 hour. Removal of platinum and palladium were over 99% while only 15% of the rhodium was recovered. This resulted in PGM loadings of around 4 500 g/t. Due to this relatively low loadings and incomplete removal of rhodium no further tests were conducted.

Ion reduction and ion exchange resins were then screened for their ability to recover the PGMs from solution. Resin was added to the leach solutions in 100 ml increments with the final addition more than 30 times the recommended requirement. Solutions were analysed after 24 hours and it was found that the best reduction resin removed up to 95% of platinum and 50% of rhodium while the best ion exchange resin removed up to 99% of platinum and 80% of rhodium. It was concluded that resins can rapidly remove PGMs from solution but that the incomplete removal thereof was still a concern.

The direct electrolysis of PGMs from solution was also investigated. While it was found that electrolysis could remove PGMs from solution, the deposition rate decreased as the PGM concentrations decreased and thus many metal impurities co-deposited. Current efficiencies of less than 1% were calculated while only 50% of the PGMs were removed after several days of electrolysis. It was thus concluded from the long time and incomplete recovery that electrolysis was not the best method of recovery.

The final method of recovery that was tested was the thermal decomposition of the cyanide complexes. It was observed that at temperatures higher than 160°C the metal concentrations in solution, predominantly palladium, decreased rapidly. This suggested that complexes formed at lower temperatures were unstable at these high temperatures and that it decomposed to precipitate to their respective metals from solution.

Tests were subsequently conducted in which the leach solutions were reheated in a pressure vessel to even higher temperatures, above 200°C for 1 hour. The results from these tests are presented in Table 3.2. PGM recoveries in the range of 99% were achieved by heating the solution to 250°C for 1 hour while recoveries of more than 99.8% were always achieved by increasing the temperature to 275°C.

Tests further showed that platinum was the most difficult to remove from solution, suggesting that the platinum cyanide complexes were the most thermally stable of the PGMs under these conditions.

Table 3.2: *Effect of temperature for 1 hour on the removal of PGMs from cyanide leach solutions (After Desmond et al., 1991)*

Temp	*Pressure*	*Partial analysis (mg/L)*		
°C	*psig*	***Pt***	***Pd***	***Rh***
Feed		145	42	14
200	210	146	1.4	12
225	350	102	0.4	5.4
250	600	0.2	<0.1	<0.1
275	900	0.4	<0.1	<0.1

Another benefit of this process is the fact that almost complete destruction of the total cyanide was achieved, paving the way for safe disposal of the barren solution from a cyanide perspective. The cyanide concentration was observed to be below 0.2 mg/L at the point where the PGMs were removed to less than 1 mg/L in solution.

From these studies it were concluded that the best method for PGM recovery from cyanide leach solutions is the thermal decomposition or thermal hydrolysis of the cyanide complexes in an autoclave, yielding a fine black powder precipitate containing mostly metallic PGMs.

3.9 Summary

From the literature reviewed above and their respective findings it can be seen that the recovery of platinum group metals from low grade Platreef ore by means of cyanidation looks promising. The extraction of the PGMs via cyanide leaching has been shown to be effective, while the recovery of the metals from the resulting solution has also been shown to be possible with the use of activated carbon, ion exchange resins and thermal hydrolysis.

All these studies were, however, performed at a feasibility stage and extensive research into each of the above mentioned technique and technologies would still be required to ensure that the methods are well understood to allow for the overall process development and ultimately the commercialisation of these technologies.

The following chapter focuses on putting all of these process options and alternatives together to form overall process flowsheets for recovering the PGMs from solution.

Chapter 4

Flowsheet Development

In this chapter all the process options for the recovery of PGMs from dilute cyanide leach solutions, as identified in Chapters 2 and 3, with specific application to the low grade Platreef ore are discussed. Different flowsheets, combining these process options, are then developed for the overall extraction and recovery of the PGMs from Platreef ore. These flowsheets were used in the subsequent simulation and comparison presented in the next few chapters.

4.1 Process Feed Streams and Composition

From the work conducted by Mwase (2009) and Mwase *et al*. (2012*a*, 2012*b*) and expert opinions on heap leaching by Peterson (2012), a two stage cyanide heap leach was proposed for the extraction of the PGMs from low grade Platreef ore. This cyanide heap leach is part of a process that includes the following steps (schematically represented by Figure 4.1):

- The run of mine ore is screened into coarse and fine ore.
- The majority of the base metals are recovered by bioleaching of the coarse ore, and treatment of the fine ore in a conventional concentrator plant.
- The spent heap is recovered, washed, neutralized and screened to remove fines which, together with the concentrator concentrate, are treated in an ammonia leach plant to further recover some of the base metals.
- The ammonia leach solution and the bioleach solution are sent to a base metal recovery plant and subsequent base metal refinery.
- The resulting ammonia leach and concentrator tailings as well as the coarse spent heap are agglomerated and stacked onto pads for the subsequent cyanide heap leach.
- It was proposed that the cyanide heap leaching be performed in two stages, as the rate of palladium and gold dissolution is superior to that of the platinum.

- A 20 day leach is thus conducted in which 20% of the platinum and the majority of the palladium and gold are recovered.
- This is followed by a 100 day leach to recover a further 35% of the platinum.

This process allows for the recovery of the majority of the base metals prior to the cyanide leach to ensure that more pure PGM leach solutions are produced.

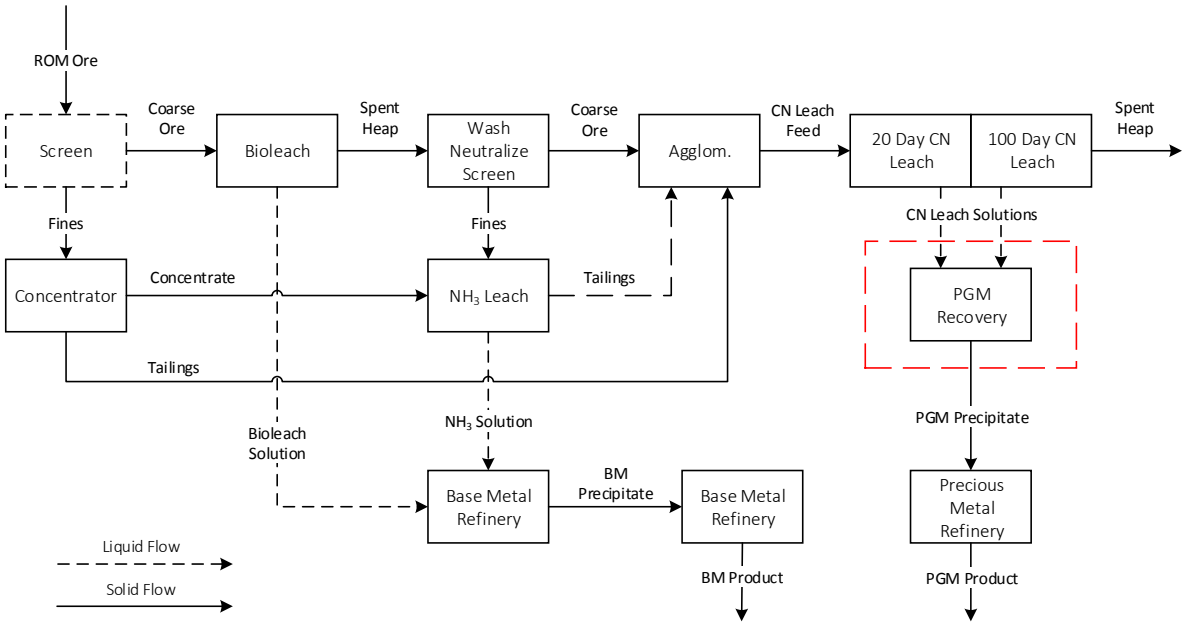

Figure 4.1: *Block flow diagram for the recovery of base and precious metals from Platreef ore*

Two leach solutions are produced from the cyanide heap leaching step and these solutions need to be treated in order to recover the PGMs that they contain. The expected flowrates and compositions of these solutions, as obtained from a mass balance over the entire process, for an assumed 240 000 t/m mining operation, are presented in Table 4.1.

Table 4.1: *Pregnant Cyanide Leach Solutions Composition and Flowrate (Mwase & Peterson, 2012)*

Primary PLS (20 Day Leach)			*Secondary PLS (100 Day Leach)*		
Flow	58.97	t/h	Flow	294.39	t/h
Cu	352.44	mg/L	Cu	52.52	mg/L
Ni	267.22	mg/L	Ni	65.98	mg/L
Zn	7.99	mg/L	Zn	2.13	mg/L
Fe	0.52	mg/L	Fe	0.21	mg/L
Pt	2.55	mg/L	Pt	0.89	mg/L
Pd	13.43	mg/L	Pd	-	
Au	1.91	mg/L	Au	-	

The objective of this project is thus to find the optimal process to recover the PGMs from these cyanide leach solutions, depicted by the dotted box in Figure 4.1, to produce a valuable, high-grade PGM precipitate or a product that can be directly fed to a precious metal refinery. The data presented in Table 4.1 were used in the development of the different flowsheet alternatives, and served as the feed streams to these processes.

4.2 Process Options and Alternatives

The requirement is to develop a process that is economically viable and can produce a PGM product that is either suitable to be processed as a direct feed to a precious metal refinery (PMR) or sold directly as a valuable, high-grade product (toll refining). Figure 4.2 shows all the process options and alternatives that could be used to accomplish this goal, as identified from the preceding chapters, and is discussed further below.

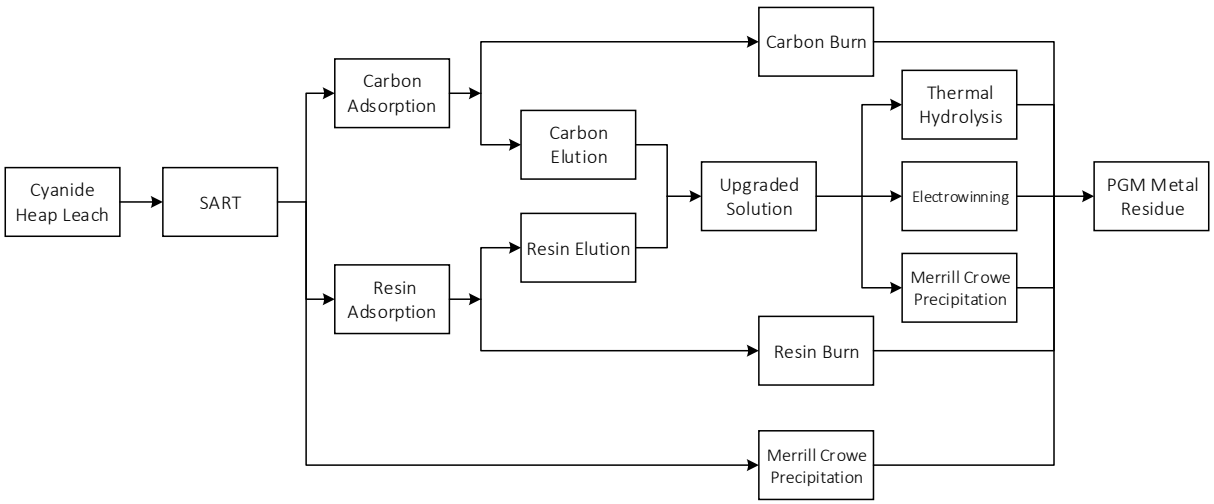

Figure 4.2: *Process options for the recovery of PGMs from low grade ores by means of cyanidation.*

4.2.1 Primary Extraction

The objective of this project is to find the optimum process to recover precious metals from dilute cyanide solutions, as produced by a cyanide heap leach operation. Hence the first step in the process is the subsequent cyanide heap leaching of low grade ore. Here the PGMs are extracted from the ore by using a cyanide solution and the resulting pregnant liquor is collected for downstream processing. This process has been shown to be viable either for the treatment of low grade concentrate or low grade ore (Mwase, 2009; Mwase *et al*., 2012*a*, 2012*b*).

4.2.2 Base Metal Removal

The mineralogy of the Platreef ore is such that extensive precious metal mineralization occurs with mainly copper, nickel, iron and zinc (Bryson, 2008). These base metals are readily leached during cyanide heap leaching (Marsden & House, 2006; Mwase, 2009) and as a result their concentrations in the pregnant liquor are very high (refer to Table 4.1), even though high base metal extractions are achieved during the preceding bioleaching of the ore.

This can have detrimental effects on the downstream processes such as adsorption onto activated carbon and ion exchange resins, as the base metals compete directly with the precious metals for active sites (Nicol *et al*., 1984*c*; Adams, 1991; Sceresini, 2005; Marsden & House, 2006; Dai & Breuer, 2009; Dai *et al*., 2010; Mpinga, 2012; Snyders *et al*., 2013). In order to lower these base metal concentrations in the leach solutions, it is proposed that the solutions be treated in a SART plant.

It has been shown that with the use of a SART plant up to 99% of base metals in such a cyanide solution can be recovered with minimal losses of precious metals (Botz *et al*., 2011). The base metals are recovered as a sulphide precipitate that can, together with a gypsum precipitate, be sold as valuable by-products. The cyanide associated with the base metals is regenerated as free cyanide in the process and can be recycled back to the heap leach for re-use (Marsden & House, 2006; Estay *et al*., 2012). Cyanide is a major contributor to the operating costs in gold heap leaching operations and thus great savings can be expected in this regard (Fleming, 1992; Marsden & House, 2006; Stewart & Kappes, 2012).

The precipitation of the base metals and the regeneration of cyanide would furthermore result in an increase of the cyanide to copper mole ratio. This would in turn lead to the formation of the copper complex $Cu(CN)_4^{3-}$, which is the least adsorbed of the copper cyanide complexes during carbon adsorption and it would therefore hold numerous benefits for the subsequent downstream adsorption and elution circuits (Adams, 1991; Boshoff, 1994; Lu *et al*., 2002).

The revenue of the by-products together with the credit for recycling the free cyanide to the heap leach can cover a great amount of the operating expenses. The SART plant at Maricunga gold mine in Chile, treating a leach solution of 750 m^3/h containing 360 mg/L Cu had a payback period of about 2 years. The payback period of a specific application would, however, depend on the copper content of the leach solution (Ford *et al*., 2008).

4.2.3 Secondary Extraction

Several options exist for the treatment of the resulting base metal free leach solutions, either adsorption of the PGMs onto activated carbon or ion exchange resins, precipitation of the PGMs via the Merrill Crowe process or thermal decomposition of the PGM cyanide complexes in an autoclave.

Extraction of platinum, palladium and gold via activated carbon has been shown to be effective (Desmond *et al*., 1991; Aguilar *et al*., 1997; Mpinga, 2012; Snyders *et al*., 2013) as well as the extraction via ion exchanges resins (Schoeman 2012, Schoeman *et al*., 2012).

The Merrill Crowe zinc cementation process has, however, been shown to be ineffective in the direct precipitation of PGMs from such dilute solutions (Desmond *et al*., 1991; Mpinga *et al*., 2013), while thermal hydrolysis as a direct treatment method of the cyanide leach liquor has been deemed uneconomical and ineffective due to the vast amounts of pregnant leach solution to be treated and the limited application to high strength solutions (Desmond *et al*., 1991; Kuczynski *et al*., 1992).

Adsorption onto activated carbon and ion exchange resins were thus the main process routes identified and investigated further. Counter-current extraction of the precious metals based on the carbon-in-solution and resin-in-solution processes were thus identified for the extraction of the precious metals from the cyanide leach liquors.

4.2.4 Upgrading Phase

The loaded carbon or resin needs to undergo treatment to recover the PGMs. This can either be done by burning the carbon or resin to ash, recovering the PGMs in a melted phase or eluting the carbon or resin to produce an upgraded solution containing the adsorbed PGMs.

The environmental issues regarding the option of burning the carbon or resin to ash could be a problem due to the disposal of the ash residue and exhaust gasses that are generated in the process, such as the potential of hydrogen cyanide generation and carbon and sulphur emissions. Furthermore the costs associated with replacing the carbon and resin in the adsorption stages would be significant and from a preliminary cost analysis it was found to render this option uneconomical compared with elution, especially in the case of the more expensive resin.

As the base metals in solution are strongly loaded onto both carbon and resin, the recovered melted phase would contain all of the adsorbed species and the subsequent grade of the product is expected to be poor with regards to the PGMs.

Both the elution of activated carbon (Snyders, 2012; Snyders *et al*., 2013) and ion exchange resins (Schoeman, 2012; Schoeman *et al*., 2012) loaded with PGMs have been shown to be effective. Acid washing has further been shown to separate the majority of the base metals from the PGMs, resulting in a valuable, high-grade precipitate when the metals are recovered from the eluate (Davidson & Schmidt, 1986; Conradie *et al*., 1994; Fisher & LaBrooy, 1997; Marsden & House, 2006).

Methods for the regeneration of carbon and resin exists, and are used industrially, that allow for the adsorbent to be recycled back to the adsorption section for re-use, eliminating the cost of replacing the adsorbent after each cycle, as would be the case if it was burnt (Fleming & Cromberge, 1984*a*, 1984*c*; La Brooy *et al*., 1984; Stange, 1999; Kotze *et al*., 2005).

Taking all of the above into consideration, the options of carbon and resin burn were discarded as higher product quality and cost savings were expected if elution were to be employed. Recovery of the PGMs from the loaded activated carbon or ion exchange resins was thus selected to be accomplished via stripping of the adsorbent in an elution circuit.

4.2.5 Final Metal Recovery

The concentrated solution produced by the elution of the activated carbon or ion exchange resin needs further treatment to recover the PGMs as a valuable product to be refined further or sold directly. This could be accomplished either via the direct electrowinning of the solutions, thermal decomposition of the cyanide complexes in an autoclave, precipitation via the Merrill Crowe process or a combination of the three process options.

As mentioned earlier, current research (Mpinga *et al*., 2013) indicated that the zinc cementation of dilute and concentrated PGM solutions are ineffective, thus leaving direct electrowinning and thermal hydrolysis as the only recovery processes.

Electrowinning of gold CIP plant eluates is a well understood and well implemented process that is used extensively in recovering gold as a high grade product at recoveries in excess of 99% (Marsden & House, 2006; van Tonder & Sole, 2006).

Electrowinning of gold RIP plant eluates has also been shown to be effective with recoveries in excess of 99% in a treatment time of less than 5 hours (Conradie *et al*., 1994). The behaviour of platinum and palladium when subjected to electrowinning is, however, unknown at present. Due to this, the direct electrowinning of the eluates that still contains high amounts of platinum and palladium was not considered.

Thermal hydrolysis has been shown to be very effective in recovering platinum and palladium from cyanide solutions, while under certain condition the recovery of gold can be limited to yield a precipitate containing around 70% metallic PGMs at recoveries in excess of 99% (Adams, 1995; Desmond *et al*., 1991; Kuczynski *et al*., 1992;). This process could allow for the separation of the PGMs from gold and allow for the direct electrowinning of the resulting solution to recover the gold.

A joint recovery process, comprising thermal hydrolysis followed by electrowinning, was thus identified for treating plant eluates containing PGMs and gold, while direct electrowinning can be used for solutions containing mainly gold and thermal hydrolyses for treating solutions containing mainly PGMs. This would yield metal residues that can either be fed to a precious metal refinery (PMR) for further refining or sold as valuable, high-grade products.

A benefit of using thermal hydrolysis as the primary recovery method is that almost complete destruction of the total cyanide is achieved during the course of the process (Desmond *et al*., 1991; Kuczynski *et al*., 1992). This paves the way for safe disposal of the treated solutions from a cyanide perspective.

4.3 Flowsheet Alternatives

Taking the above options into consideration, different flowsheets for the recovery of precious metals from dilute cyanide solutions have been developed. These flowsheets were classified according to the main process route that was followed, i.e. either adsorption onto activated carbon or adsorption onto ion exchange resins. These flowsheets, along with the variations of the above mentioned processes and their application in each, are discussed in detail in the next section.

4.3.1 Activated Carbon Flowsheets (Carbon-in-Solution)

For the treatment of the cyanide leach liquors via adsorption onto activated carbon, two different flowsheet alternatives were developed. These flowsheets follow the general carbon adsorption process route, as outlined by Stange (1999), and are discussed in detail below.

4.3.1.1 Flowsheet Alternative 1 (CIS 1)

The first CIS flowsheet (CIS 1) follows the same general route that is used in the extraction of gold by means of activated carbon. A block flow diagram of the process, indicating the main process route, is presented in Figure 4.3, while a more detailed process flow diagram can be found in Appendix A.

It was assumed that, under steady state operating conditions, the two cyanide leach solutions could be combined to form one solution to be treated. The solution is first treated in a SART plant to remove the excessive amounts of base metals and to free up the associated cyanide for recycle to the heap leach stage.

The next step is the counter-current extraction of the value material in a carbon adsorption circuit. The resulting barren solution, containing the free cyanide generated during SART, can then be partially recycled back to the heap leach to reduce the cyanide consumption. The precious metals still left in this stream are believed to further enhance the performance of the leaching step. This has, however, not been tested and is unsure at present.

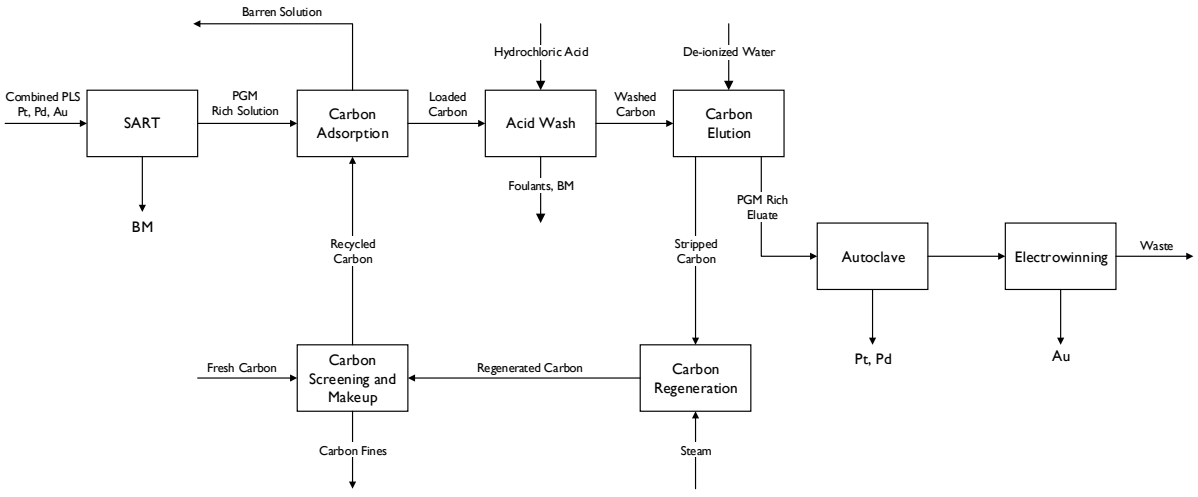

Figure 4.3: Block Flow Diagram of Flowsheet Alternative 1 (CIS 1)

The loaded carbon is then first screened to remove carbon fines where-after it undergoes an acid wash treatment. A dilute hydrochloric acid solution is used, which removes the organic foulants from the carbon as well as about 50% of the nickel and zinc that is likely to be co-extracted in the adsorption circuit (Davidson & Duncan, 1977; Fisher & LaBrooy, 1997; Marsden & House, 2006). The acid washing step was selected to be performed prior to elution due to the multiple benefits that it may hold for the subsequent elution and regeneration circuit performance, as discussed in section 2.5.4.

The acid washed carbon is then transferred to the elution column where it is first pre-treated with a sodium cyanide, sodium hydroxide solution and is then eluted by means of the AARL procedure to produce a concentrated solution rich in precious metals. Simultaneously elution of platinum, palladium and gold can be achieved with this method (Eksteen *et al*., 2013; Snyders *et al*., 2013). The stripped carbon is then transferred to the regeneration circuit while the pregnant eluate is sent to the downstream recovery circuit.

As the eluate from the elution stage will contain high amounts of both PGMs and gold, the joint recovery process, i.e. thermal hydrolysis followed by electrowinning is employed. The eluate is first sent to an autoclave, where under high temperatures and pressure, the PGMs are recovered as a valuable high grade precipitate. The resulting solution, still rich in gold, is then circulated through electrowinning cells to recover the gold.

The spent eluate, believed to be stripped from almost all metals, is then sent to the tailings dam as it was assumed to be environmentally benign from a cyanide perspective, due to the destruction thereof during the thermal hydrolysis step (Desmond *et al*., 1991).

The stripped carbon is then regenerated in a rotary kiln, at about 750°C in the presence of steam, to restore the activity of the carbon to allow for the recycle and reuse in the adsorption circuit. The warm activated carbon is first quenched in a water quench tank and then screened to remove fines generated during the regeneration cycle. Makeup carbon is added to the oversized carbon as to ensure that the required amount of carbon is recycled back to the adsorption circuit to achieve optimum performance.

This process would thus allow for the simultaneous recovery of platinum, palladium and gold and is believed to produce valuable high grade PGM and gold products that can either be toll refined or fed to a precious metal refinery.

4.3.1.2 Flowsheet Alternative 2 (CIS 2)

The second carbon flowsheet alternative (CIS 2) is generally the same as the previous flowsheet. However, the leach solutions from the two heap leaching stages are not combined and are therefore subsequently treated in separate plants.

The process route for the primary leach solution would thus be exactly the same as the previous flowsheet alternative (CIS 1), as it contains both platinum, palladium and gold, while the secondary leach solution process route would only differ in the sense that an electrowinning circuit is not required as only platinum would be present in the feed to this process. A block flow diagram of this proposed process is presented in Figure 4.4, while a more detailed process flow diagram can be found in Appendix A.

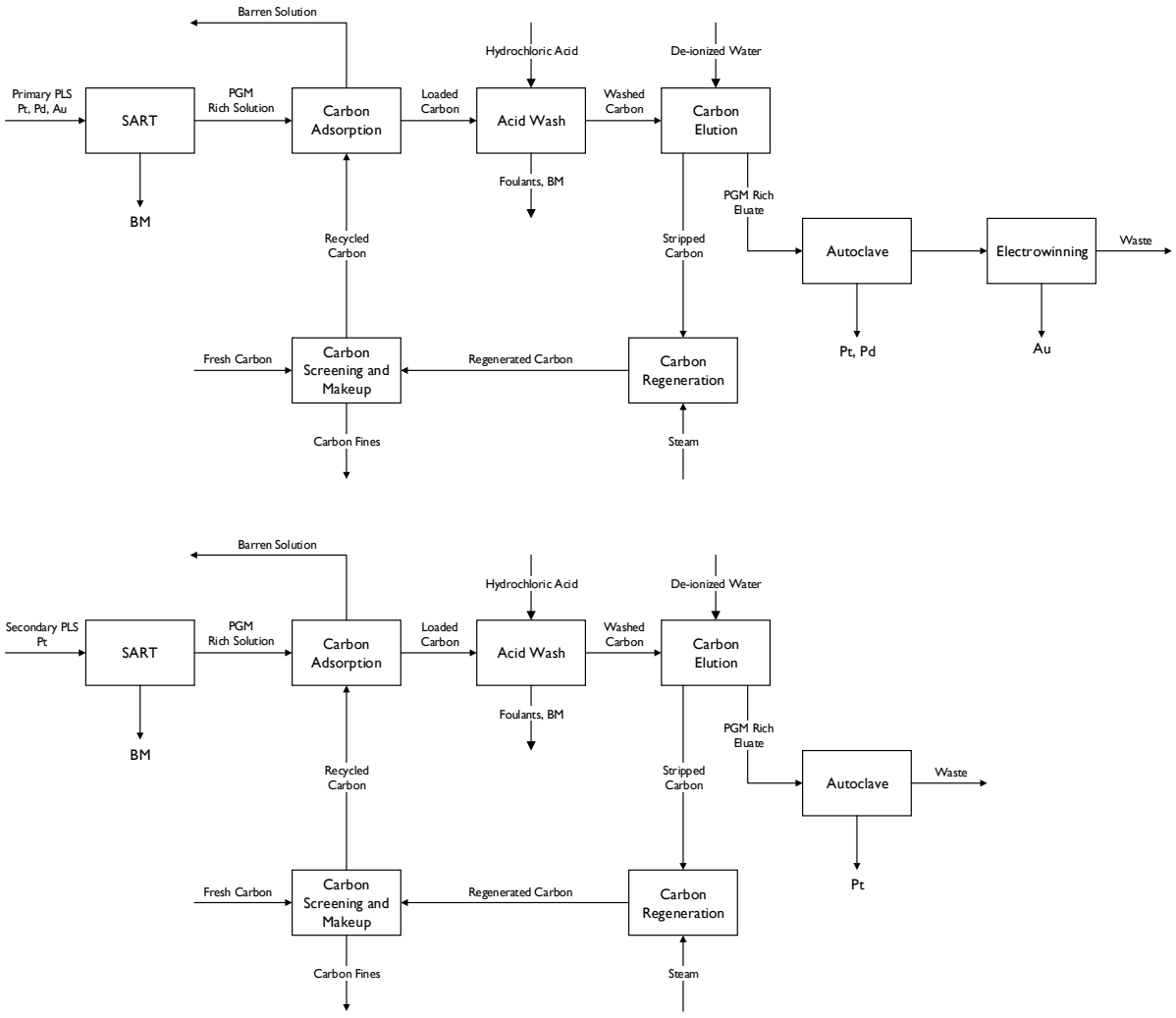

Figure 4.4: *Block Flow Diagram of Flowsheet Alternative 2 (CIS 2)*

4.3.2 Ion Exchange Resin Flowsheets (Resin-in-Solution)

For the treatment of the cyanide leach liquors via adsorption onto ion exchange resins, five different flowsheet alternatives were developed. The reason for the increased number of resin flowsheets when compared to the activated carbon flowsheets is that there are multiple resins and eluting agents that can be used in the application of ion exchange resins to cyanide leach liquors (refer to sections 3.5 and 3.6).

4.3.2.1 Flowsheet Alternative 3 (RIS 1)

The first RIS flowsheet (RIS 1) is the most simplistic of all. Only one resin, Amberlite PWA 5, is used as it is effective in adsorbing platinum, palladium and gold from cyanide leach solutions (Schoeman, 2012; Schoeman *et al*., 2012). Furthermore only one of the possible eluting agents, zinc cyanide, is used, as it is relatively effective in stripping all metals from this resin (Schoeman, 2012).

The general process route of this process is similar to that of the first carbon flowsheet (CIS 1) and is discussed in further detail below. The process is presented in Figure 4.5, while a more detailed process flow diagram can be found in Appendix A.

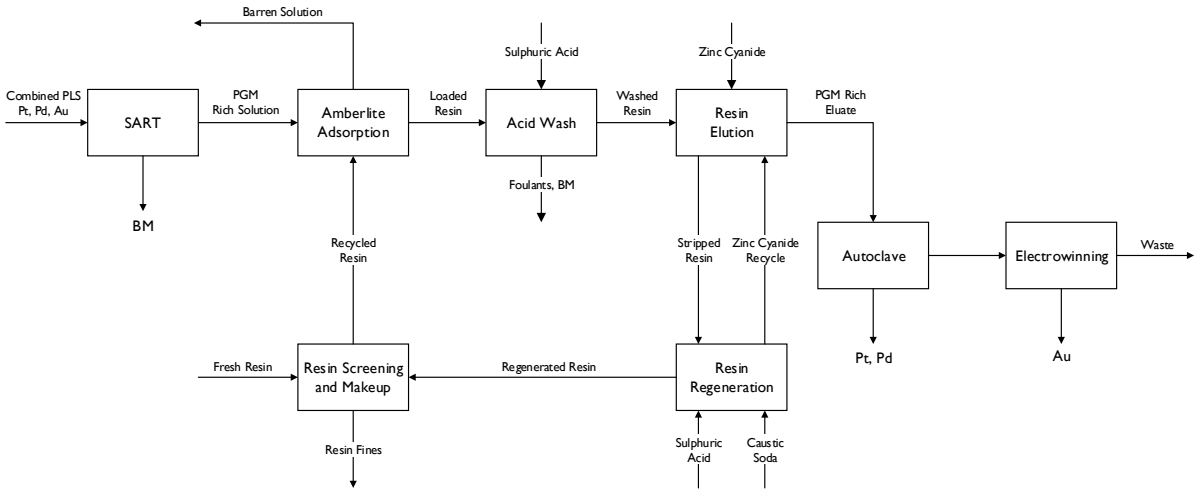

Figure 4.5: Block Flow Diagram of Flowsheet Alternative 3 (RIS 1)

In this process the leach solutions are again combined to form one solution containing platinum, palladium and gold. The solution is first treated in a SART plant where-after it is fed to a counter-current resin (Amberlite PWA 5) adsorption circuit. The barren solution can then be partially recycled back to the heap leach stage while the loaded resin is transferred to the downstream elution circuit.

The loaded resin, first screened to remove any fines, is transferred to an acid washing step where the majority of the loaded nickel and zinc is removed by washing the resin with a dilute sulphuric acid solution. Conradie *et al*. (1994) have reported that almost complete removal of the nickel and zinc can be achieved.

The acid washed resin is then transferred to the elution column where it is eluted with a zinc cyanide solution. The eluate, rich in both PGMs and gold, is fed to the recovery circuit where the PGMs are recovered in an autoclave and the gold by means of electrowinning, similar to that of the CIS 1 flowsheet option. The barren solution is then sent to the tailings dam.

The stripped resin needs to go through a regeneration step, as the zinc cyanide used in the elution cycle is strongly adsorbed onto the resin, reducing its activity. If recycled back to the adsorption circuit untreated, a significant decrease in the performance of the adsorption circuit will be observed (Fleming & Cromberge, 1984*a*, 1984*c*; Kotze *et al*., 2005).

The regeneration of the resin is performed using sulphuric acid, which strips the zinc cyanide from the resin. The resulting solution is fed to a neutralization stage where the zinc cyanide is recovered and recycled for reuse in the elution circuit. Prior to the next elution, the zinc cyanide concentration must be restored to the required concentration, by the addition of zinc sulphate and sodium cyanide.

The regenerated resin is then screened, removing any fines generated during elution and regeneration, with make-up resin being added prior to recycling it back to the adsorption circuit.

This process would allow for the simultaneous recovery of platinum, palladium and gold and is believed to produce valuable high grade PGM and gold products that can either be toll refined or fed to a precious metal refinery.

4.3.2.2 Flowsheet Alternative 4 (RIS 2)

The second RIS flowsheet alternative (RIS 2) is generally the same as the first (RIS 1). However, the resulting leach solutions from the cyanide heap leaching stage are not combined and are therefore subsequently treated in separate processing plants.

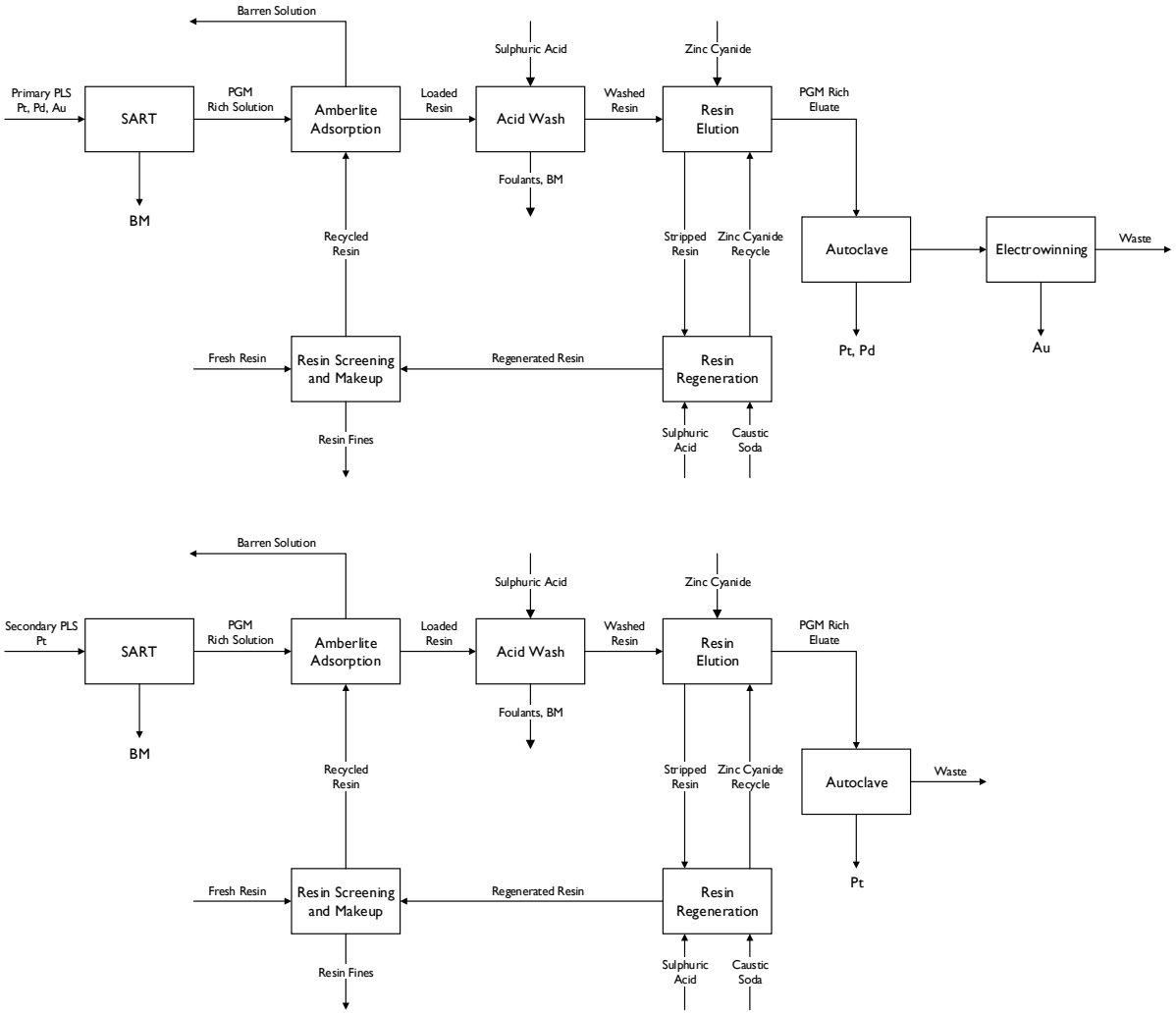

Figure 4.6: *Block Flow Diagram of Flowsheet Alternative 4 (RIS 2)*

The process route for the primary leach solution would thus be exactly the same as the previous flowsheet (RIS 1), as it contains both platinum, palladium and gold, while the secondary leach solution process would only differ in the sense that an electrowinning circuit is not required as only platinum are believed to be present in the feed to this process.

A block flow diagram of the process is presented in Figure 4.6, while a more detailed process flow diagram can be found in Appendix A.

4.3.2.3 Flowsheet Alternative 5 (RIS 3)

The third RIS flowsheet (RIS 3) is generally the same as the first (RIS 1) and only differs in the sense that a sequential elution procedure is employed.

After the adsorption section, the resin is first eluted using an acidic thiourea solution. This eluting agent is very effective in stripping the gold and palladium loaded onto the Amberlite PWA 5 resin, and has been shown to have improved gold elution efficiencies when compared to the elution of strong base resins with zinc cyanide (Fleming & Cromberge, 1984*a*, 1984*c*; Kotze *et al*., 2005; Schoeman, 2012). Acidic thiourea has, however, been found to be ineffective in recovering the loaded platinum from this resin (Schoeman, 2012; Schoeman *et al*., 2012). Reasons for this were not provided and this phenomenon could prove to be problematic if the adsorbed platinum is altered so that normal elution procedures cannot be employed. It was, however, assumed that the loaded platinum is unaffected to allow for effective elution thereof. This allows for the separation of the value material as the eluted resin, still highly loaded with platinum, is subjected to subsequent zinc cyanide elution that strips the resin of the loaded platinum and the palladium and gold that was not eluted in the first elution cycle.

Two separate solutions, one containing the majority of the palladium and gold and one containing the platinum, are thus produced. The eluate, rich in both palladium and gold, is sent to an autoclave to recover the palladium and a subsequent electrowinning stage to recover the gold, with the barren solution sent to the tailings dam.

The platinum is recovered by sending the secondary eluate to an autoclave. It is assumed that an electrowinning circuit is not required as the gold content of this eluate is expected to be below the economic cut-off point. The barren solution is then sent to the tailings dam.

As zinc cyanide has been used in the elution circuit regeneration of the resin is required prior to it being recycled back to the adsorption circuit. The resin regeneration is the same as that employed in the first resin flowsheet (RIS 1).

A block flow diagram of the process, as discussed above, is presented in Figure 4.7, while a more detailed process flow diagram can be found in Appendix A. This process would thus enable the separation of all three value materials, platinum, palladium and gold, and is expected to recover each metal in separate valuable high grade products.

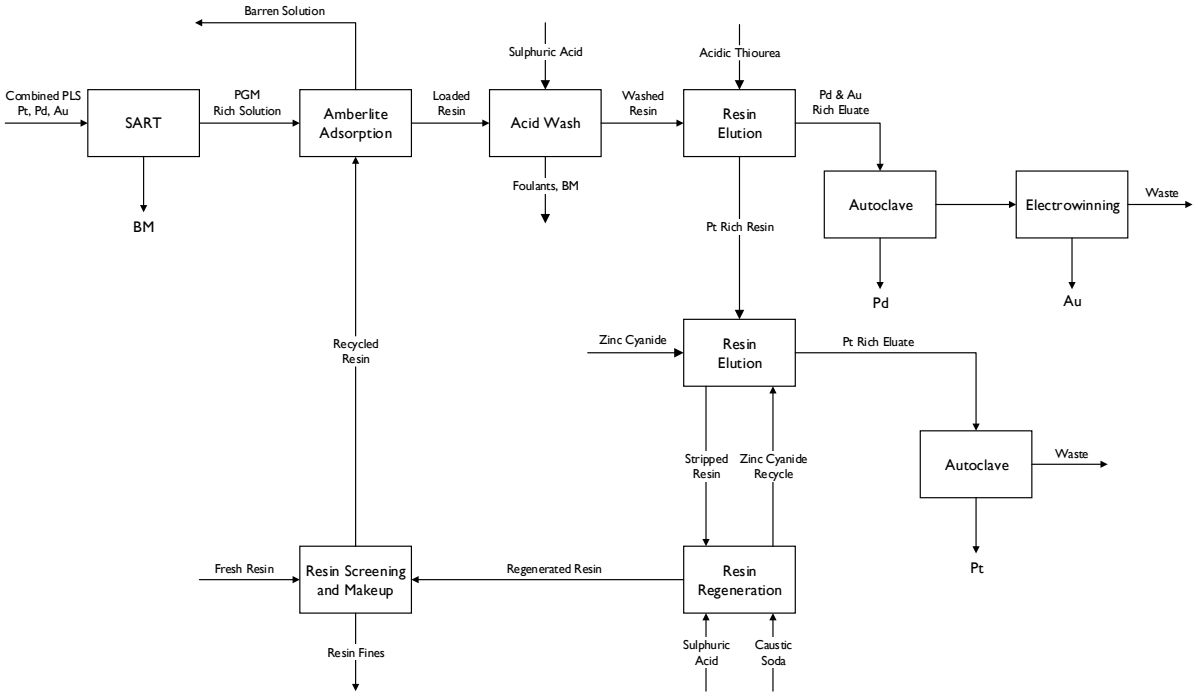

Figure 4.7: *Block Flow Diagram of Flowsheet Alternative 5 (RIS 3)*

4.3.2.4 Flowsheet Alternative 6 (RIS 4)

The fourth resin flowsheet (RIS 4) is a combination of the first (RIS 1) and third (RIS 3) resin flowsheet alternatives as the leach solutions are proposed to be treated in separate processing plants.

In this process the primary solution, containing platinum, palladium and gold, is treated via the same process route as discussed for the third RIS flowsheet (RIS 3) above, while the secondary solution, only containing platinum, are treated via the same process route as discussed for the first RIS flowsheet alternative (RIS 1).

Two separate processing plants are thus required for treating the individual cyanide leach solutions. A block flow diagram of this process is presented in Figure 4.8, while a more detailed process flow diagram can be found in Appendix A.

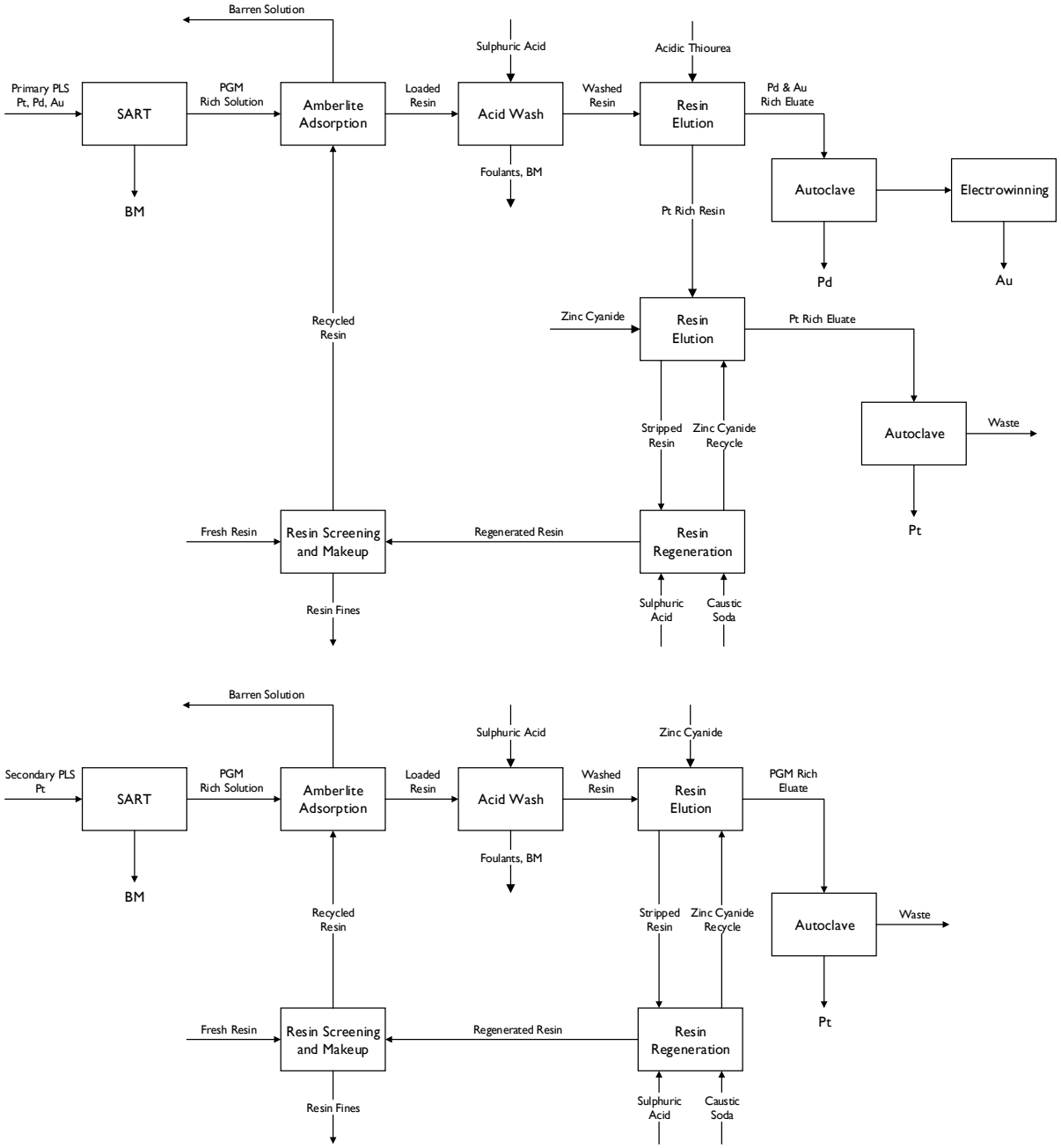

Figure 4.8: *Block Flow Diagram of Flowsheet Alternative 6 (RIS 4)*

4.3.2.5 Flowsheet Alternative 7 (RIS 5)

The last of the RIS flowsheets (RIS 5), and the most complex, utilises both of the available resins (Amberlite PWA 5 and Minix) and both of the eluting agents (Acidic thiourea and zinc cyanide). A block flow diagram of this process is presented in Figure 4.9, while a more detailed process flow diagram can be found in Appendix A.

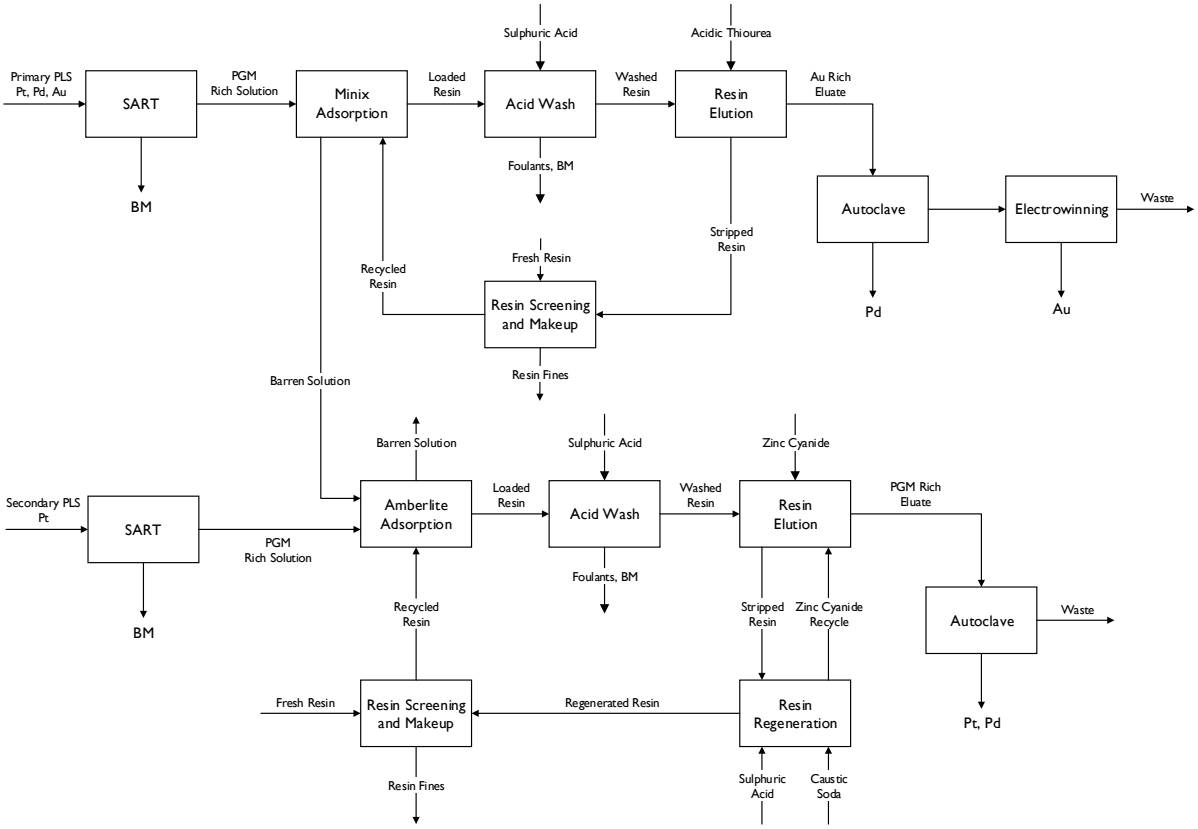

Figure 4.9: *Block Flow Diagram of Flowsheet Alternative 7 (RIS 5)*

In this process the two leach solutions are treated separately. The primary leach solution is first treated in a SART plant where-after it is fed to a counter-current adsorption circuit equipped with the Minix gold selective resin. The Minix resin has been shown to be highly selective towards gold while loading very low amounts of both platinum and palladium (Schoeman, 2012; Schoeman *et al*., 2012). This would allow for the extraction of gold and minimal amounts of the other base and precious metals, separating the gold from the other value minerals.

The loaded Minix resin is then screened, to remove any fines generated during adsorption, and transferred to the elution circuit where it is first acid washed, to remove nickel and zinc loaded during adsorption, and then eluted with an acidic thiourea solution.

Acidic thiourea has been proposed as the preferred eluting agent for the elution of gold from selective strong base resins (Fleming & Cromberge, 1984*c*; Kotze *et al*., 2005) and has been shown to have fast elution kinetics, capable of achieving almost complete gold elution (Fleming & Cromberge, 1984*a*, 1984*c*; Kotze *et al*., 2005; Schoeman, 2012; Schoeman *et al*., 2012). As mainly gold is loaded on the resin, a second elution step is not required.

The loaded palladium would be stripped from the resin while the low amounts of the loaded platinum would remain on the resin, as it has been reported that acidic thiourea does not elute platinum loaded on the Minix resin (Schoeman, 2012; Schoeman *et al*., 2012).

It was deemed that the loaded platinum, expected to be very low, would not warrant a subsequent zinc cyanide elution cycle and recovery circuit, as the capital and operating costs thereof are believed to be higher than the loaded platinum value. It was thus decided that this loaded platinum would be treated in the same manner as the locked up material in the adsorption circuits, when the economic analysis of the flowsheets are performed.

The resulting eluate, rich in gold, is then sent to an autoclave to recover the palladium and base metals that were co-extracted during adsorption and elution. The resulting solution, rich in gold is sent to an electrowinning circuit to recover the gold, while the barren solution is sent to the tailings dam.

The stripped Minix resin is recycled back to the adsorption circuit without the need of a regeneration step. Ion exchange resins eluted with acidic thiourea does not require any regeneration, as the resin is restored back to its chlorinated form during elution, as a result of the hydrochloric content of the eluting agent (Fleming & Cromberge, 1984*a*, 1984*b*, 1984*c*; Kotze *et al*., 2005).

The resulting solution leaving the Minix adsorption circuit, believed to contain almost all of the platinum and palladium, is then combined with the secondary leach solution (containing platinum) and treated via the same basic process route as the first RIS flowsheet alternative (RIS 1). Thus an adsorption circuit utilising the Amberlite PWA 5 resin and the subsequent elution with zinc cyanide, with final metal recovery performed in an autoclave.

This process would allow for the selective adsorption of firstly gold, followed by platinum and palladium, with the use of two separate adsorption circuits comprising of either the gold selective Minix resin or the PGM selective Amberlite PWA 5 resin. This negates the need for the sequential elution procedures as proposed for the RIS 3 and RIS 4 flowsheet options.

4.3.3 Flowsheet Development Summary

Seven different flowsheet alternatives were developed based on the information presented in Chapters 2 and 3. Two carbon flowsheets, analogous to the well-known carbon-in-pulp process, and five different resin flowsheets were developed and discussed. These flowsheets were then to be simulated, with economic analysis preformed for each, to allow for the flowsheets to be optimized to ultimately propose the most economically favourable treatment option for recovering precious metals from cyanide leach solutions.

The following chapter focuses on the design and simulation decisions that were made in order to simulate the different flowsheet alternatives.

Chapter 5

Basis of Design

This chapter focuses on the design decisions and assumptions that were made in order to simulate each of the unit operations in the developed flowsheets.

5.1 Assumptions and Operating Conditions

As this is a concept phase study, only the base and precious metals listed in Table 4.1 were tracked throughout the process, due to insufficient data available on the behaviour of the gangue minerals such as silica, aluminium, calcium and magnesium.

5.1.1 SART

5.1.1.1 Sulfurization and Acidification

The first step in a typical SART process is the adjustment of the pH. This was assumed to be performed with a 33% sulphuric acid solution (H_2SO_4) at the stoichiometric requirement based on the free cyanide concentration, converting the cyanide into hydrogen cyanide (HCN). Sodium hydrosulphide ($NaHS$) is then added at 120% of the stoichiometric requirement, based on the total base metal content. This value has been shown to yield the best precipitation results with copper recoveries in excess of 99% (Ford *et al*., 2008; Botz *et al*., 2011).

5.1.1.2 Base Metal Recovery

As the literature available on the SART process mainly focused on copper associated with gold ores, limited information was available on the behaviour of the other base metals during this process. As the process is based on the fact that the copper cyanide complexes are weak acid dissociable (Estay *et al*., 2012), the nickel and zinc were assumed to behave in the same manner, as their respective complexes are also weak acid dissociable (Marsden & House 2006). Ford *et al*. (2008) has mentioned that these metals are expected to behave similarly to copper.

This assumption was further justified on the basis that the stabilities of these complexes are in the same range and that they are also weak, referring to Table 5.1. Hence, 99% recovery of these metals was assumed, while no precious metals were assumed to precipitate, based on their stabilities and strength of complexes. There are, however, precious metal losses due to the solid-liquid separation step and filter cake moisture content. Losses have been reported by Ford *et al*. (2008) to amount to less than 0.5% of the total gold in solution.

Table 5.1: *Stability constants of various metal cyanide complexes at 25°C (Dionex Corporation, 2003)*

Metal Complex	*Stability Constant (log K at 25°C)*	*Type of Complex*
$Cu(CN)_4^{3-}$	23.1	Weak
$Ni(CN)_4^{2-}$	30.2	Weak
$Zn(CN)_4^{2-}$	19.6	Weak
$Fe(CN)_6^{4-}$	35.4	Strong
$Pd(CN)_4^{2-}$	42.4	Strong
$Pd(CN)_4^{2-}$	40.0	Strong
$Au(CN)_2^-$	37.0	Strong

In order to yield better precipitation settling rates, the flocculant Hychem AF 305, has been used in industry at an addition rate of 50 g/t solids (Botz *et al*., 2011). This specific flocculant and addition rate was thus assumed for use in the base metal precipitate settler.

The base metal settler underflow typically contains 12 - 16% solids (Ford *et al.*, 2008; Botz *et al*., 2011) while 70 - 90% is recycled back to the base metal precipitation reactor to enhance precipitation rates by at least 10 times that of the unconditioned raw precipitate (Estay *et al*., 2012; Stewart & Kappes, 2012). An underflow density of 15% at 80% recycle was assumed.

Two pore volumes are usually used as the filter wash water, with the base metal precipitate after filtration being reported to contain 50 - 60% moisture at a bulk density of 532 kg/m^3 on a dry basis and 1 330 kg/m^3 on a wet basis (pore volume of 1 125 kg/m^3) (Ford *et al.*, 2008; Botz *et al*., 2011).

Two pore volumes were selected to be used as the filter wash water producing a filter cake with a 50% moisture content. The resulting filter water was selected to be recycled back to the base metal precipitation reactor to ensure minimum precious metal losses.

5.1.1.3 Neutralization

The settler overflow then undergoes neutralization by the addition of lime, ($Ca(OH)_2$), to a pH of 10.5 - 11 (Barter *et al*., 2001; Estay *et al*., 2010; MacPhail *et al*., 1998). A 20% lime solution, at a stoichiometric feed rate to convert all the aqueous hydrogen cyanide (HCN_{aq}) to calcium cyanide ($Ca(CN)_2$), was assumed to be used. This cyanide is then in a form that can be recycled back to the heap leach step.

As a result of the addition of the lime and acid, a gypsum by-product is formed (Ford *et al.*, 2008; Botz *et al*., 2011; Estay *et al*., 2012; Stewart & Kappes, 2012). It was assumed that the gypsum is removed by thickening at a typical slurry density of 15% solids with the addition of flocculant at 24 g/t solids (Ford *et al.*, 2008; Botz *et al*., 2011). Furthermore 80% of this stream is recycled back to the neutralization reactor, while the gypsum precipitate is filtered and washed with 2 pore volumes to produce a filter cake containing about 20% moisture (Ford *et al.*, 2008). The resulting filter wash water is recycled back to the neutralization reactor.

5.1.1.4 Lime Scrubbing

For all reactors and thickeners it was assumed that 5% of the hydrogen cyanide is volatized, as reported by Ford *et al.* (2008), and captured for treatment in the lime scrubber system. A 5% lime solution was selected to transform the volatile hydrogen cyanide to a form ($Ca(CN)_2$) that can be recycled back to the heap leach.

5.1.2 Carbon Adsorption, Elution and Regeneration

5.1.2.1 Adsorption

Adsorption recoveries of the platinum, palladium and gold were calculated based on the kinetic modelling of the adsorption circuits, as will be discussed in Chapter 7. The recoveries were in excess of 98% for all cases.

The kinetic constants were, however, used at 80% of their respective estimated values. This was due to the fact that the batch adsorption tests did not include base metals that would have affected the rate of adsorption and the fact that fouling would play a major role in the actual operation of the plant, a factor that is not taken into consideration by the model employed (Nicol *et al*., 1984*b*; La Brooy *et al*., 1984; Staunton, 2005).

Furthermore the degree of mixing achieved during batch adsorption tests is superior to that of actual adsorption tanks (Nicol *et al*., 1984*b*). This would result in slower kinetics in plant operation (Nicol *et al*., 1984*c*). This assumption will ensure that overly optimistic results are not obtained.

Kinetics for the base metals were, however, not determined and data from adsorption tests performed by Mpinga (2012) were used for the recovery of these metals. The recoveries reported were at the optimum conditions determined by Mpinga (2012) for the extraction of platinum, palladium and gold after 2 hours of extraction. Under these conditions effective recovery of the PGMs were achieved.

Because kinetic models for the base metals were not used and also because Mpinga (2012) found that after 2 hours almost all PGMs were extracted, it was assumed that the recoveries at 2 hours would provide an accurate representation of the expected base metal recoveries as simulation results for this study found that total solution residence times were less than 2 hours for all alternatives. This assumption would thus yield worse case scenarios.

The expected recoveries for the extraction of base and precious metals by the adsorption onto activated carbon, as used in the simulations, are given in Table 5.2.

Table 5.2: *Expected recoveries of various metals during adsorption onto activated carbon*

Metal	*% Recovery*
Pt	>97†
Pd	>98†
Au	>99†
Cu	25
Ni	90
Zn	80
Fe	10

† *Calculated by model*

Carbon losses due to abrasion in CIS circuits are a major design consideration. It is required that the amount of carbon fines lost, which are loaded to significant amounts of value material, is minimized, as the value of the material lost could be significant. No models or relationship for the rate of carbon fines generation were found in literature. This is due to the high degree of difficulty in monitoring fines generation during actual plant operation (Davidson & Schoeman, 1991).

Losses during the adsorption circuit typically accounts for 50% of the total fines (Davidson & Schoeman, 1991; Fleming, 1992; Marsden & House, 2006), while losses during thermal elution and regeneration is about 5% of the total feed to the reactivation circuit (Zanitsch & Lynch, 1978). A study conducted on the carbon consumption at 18 large CIP operations found that the average carbon loss was 69 ± 59 g/t (Davidson & Schoeman, 1991).

In order to account for the carbon loss in the adsorption circuit, which is believed to be a function of the carbon inventory and residence time (Barbetti *et al.*, 2000; Stanton, 2005), a linear relationship between the carbon residence time and fines generation of 0.02% per hour was assumed. Davidson & Schoeman, (1991) has mentioned that a linear relationship between the two factors does exist. From this assumption and the losses during elution and regeneration (discussed later), total carbon losses of between 30 - 70 g/t were estimated in the simulations. This compared well to the above mentioned losses, taking note that carbon losses in CIS circuits are usually 30 - 40% less than that observed in CIP circuits (Marsden & House, 2006).

5.1.2.2 Acid Wash and Pre-treatment

Acid washing was assumed to be performed with the carbon being soaked in 1 bed volume of a dilute (3% v/v) hydrochloric acid solution at 90°C for 1 hour. This method has been proven to be the optimum and have been employed by several CIP operations (Sceresini, 1982; Fisher & LaBrooy, 1997; Boshoff, 1994; Stange, 1999; Marsden & House, 2006). During the acid washing step, nickel and zinc together with the organic foulants are stripped from the resin (Fisher & LaBrooy, 1997; Stange, 1999; Marsden & House, 2006). Nickel and zinc recoveries were assumed to be 50% each at optimum operating conditions (Fisher & LaBrooy, 1997; Marsden & House, 2006). Washing of the carbon, to remove residual acid, is performed using 1 bed volume of water.

Pre-treatment of the acid washed carbon is required to re-dissolve the material that may have been reduced to a metallic state during acid washing, and ensure effective recovery of the value material (Davidson & Schmidt, 1986; Jones *et al.*, 1988; Boshoff, 1994; Fisher & LaBrooy, 1997). Optimum pre-treatment conditions have been reported by Eksteen *et al*. (2013) for carbon loaded with base and precious metals. These conditions are 1 bed volume of a solution containing 0.55% $NaOH$ and 2% $NaCN$ at 120°C. It was assumed that no stripping of any metals occurs during this step and that there is no recycling of the acid or pre-treatment solutions during these operations.

5.1.2.3 Elution

Snyders (2011), Snyders *et al*. (2013) and Eksteen *et al*. (2013) performed extensive work on the elution of platinum, palladium and gold from activated carbon by means of the AARL elution method. Platinum and palladium recoveries were found to be in excess of 98% after 8 bed volumes (BV) of eluate at 80°C, while gold recovery was only about 50%. This is because gold elutes at higher temperatures, at around 110 - 120°C (Davidson & Duncanson, 1977; Davidson & Veronese, 1979; Davidson & Schmidt, 1986; Boshoff, 1994; van Deventer & van der Merwe, 1994). It was thus assumed that gold recovery from activated carbon loaded with platinum, palladium and gold would be at around 99% if such high elution temperatures are employed.

Snyders (2011) found that elution of carbon loaded to higher values required longer elution times. Carbon loaded with 650 g/t of platinum and palladium each, required about 6.5 BV, while carbon loaded to 3450 and 3 625 g/t required almost 10 BV to achieve the same elution efficiencies (98%), with palladium eluting slower than platinum. A linear relationship between loading and the required amount of eluate (in BV) was thus assumed.

Data from work performed by Davidson & Duncanson (1977), Davidson & Veronese (1979), Davidson & Schmidt (1986), Boshoff (1994) and Fisher & LaBrooy (1997) on the elution of gold and base metals from activated carbon were used to predict the subsequent base metal recoveries. These recoveries are typically in excess of 90%.

Elution was thus assumed to be performed according to the AARL elution method, with deionized water at 120°C at 2 BV/h. The recoveries used can be seen in Table 5.3.

Table 5.3: *Expected recoveries of various metals during elution of activated carbon*

Metal	*% Recovery*
Pt	98.5
Pd	98
Au	99
Cu	92
Ni	96
Zn	98
Fe	20

It has been assumed that there is no eluate recycle after final metal recovery, as the quality of the water would be insufficient to achieve optimum elution efficiencies.

5.1.2.4 Carbon Regeneration

Regeneration of the carbon was assumed to be performed in a rotary kiln. This is required to remove the inorganic foulants from the carbon to restore it to its original activity (Davidson & Schoeman, 1991; Fleming, 1992; Stange, 1999). A reactivation temperature of 700°C was selected, with the carbon regenerated in a fuel fired kiln in the presence of steam, to ensure minimal carbon breakage due to thermal and osmotic shock (Zanitsch & Lynch, 1978). The specific temperature was selected as it is the average temperature used in industry. The activity can be enhanced even further by regeneration above 800°C but excessive oxidation and degradation of the carbon particles occurs (Davidson & Schoeman, 1991).

Typical carbon losses observed during elution and regeneration are between 5 - 7% (Zanitsch & Lynch, 1978; Davidson & Schoeman, 1991), responsible for approximately 40 - 50% of the total carbon loss in a CIP circuits (Davidson & Schoeman, 1991; Marsden & House, 2006). A value of 6% was thus assumed as the carbon lost to fines during elution and regeneration.

5.1.3 Resin Adsorption, Elution and Regeneration

5.1.3.1 Adsorption

Adsorption recoveries of the platinum, palladium and gold were calculated based on the modelling of the adsorption circuits, as discussed in Chapter 7. The recoveries of these metals varied depending on the type of resin used, either Amberlite PWA 5 or Minix. Kinetics of platinum and palladium extraction with Minix were, however, not determined, due to the high selectivity of the Minix resin for gold, resulting in the platinum and palladium loaded during the initial stages of adsorption being rejected by the resin. Data from Schoeman (2012) were used to predict their respective recoveries.

From the work conducted by Schoeman (2012) during continuous loading tests, that closely simulate a counter-current adsorption circuit, it was found that the maximum platinum and palladium loadings onto Minix were 155 g/t and 62 g/t respectively, due to the metals being rejected as a result of the high selectivity of the resin for gold. The solutions used by Schoeman (2012) contained only 0.1 mg/L of gold, and it was assumed that this trend would be the same at the concentrations used in this study (> 0.3 mg/L). It was thus assumed that these values would be the maximum loadings achievable during simulation of the different adsorption circuits.

Data obtained during batch and continuous loading experiments performed by Schoeman (2012) were used to estimate the base metal recoveries for each resin. The expected recoveries for the extraction of the base and precious metals by Amberlite PWA 5 and Minix are given in Table 5.4.

Table 5.4: *Expected recoveries of various metals during adsorption onto ion exchange resins*

Amberlite PWA 5		*Minix*	
Metal	*% Recovery*	*Metal*	*% Recovery*
Pt	>99†	Pt	‡
Pd	>98†	Pd	‡
Au	>75†	Au	>99†
Cu	20	Cu	5
Ni	88	Ni	10
Zn	94	Zn	64
Fe	5	Fe	3

† Calculated by model

‡ Limited by Maximum Loading

Resin losses due to abrasion, physical transport and handling are significantly less compared to activated carbon (Fleming, 1992; Kotze *et al.*, 2005). This is due to the physical spherical shape and abrasive resistance of industrial produced ion exchange resins.

At the Golden Jubilee Mine where a CIP operation was replaced by an RIP operation, resin losses were found to be 6 - 10 times less than for activated carbon (8 - 10 g/t) (Fleming, 1992). It was thus assumed that the resin losses during adsorption will be 8 times less than that in the carbon adsorption circuits, i.e. 0.0025% per hour, while elution losses were assumed to be the same, at 1% of the total resin capacity. This assumption resulted in resin losses of 2 - 8 g/t, which compared well to that observed at the Golden Jubilee mine, taking in consideration that losses in RIS are about 30 - 40% less than that in RIP operations (Marsden & House, 2006).

5.1.3.2 Acid Wash

Acid washing was assumed to be performed with the resins being washed in 8 bed volumes of a 1 M sulphuric acid solution at 60°C. This method has been shown to remove nickel and zinc in excess of 98% (Conradie *et al.*, 1995). No acid recycle or neutralization was assumed during these operations.

5.1.3.3 Elution

Due to the lack of reliable elution models, data from Schoeman (2012), Schoeman *et al*. (2012) and Fleming & Cromberge (1984*c*), on the elution of base and precious metals from ion exchange resins, were used to predict the elution recoveries in the different resin elution circuits.

Schoeman (2012) provided results on the elution of base and precious metals loaded onto Amberlite PWA 5 with the use of zinc cyanide and acidic thiourea. For the zinc cyanide elution of Amberlite PWA 5, optimal conditions were found to be 0.5 M zinc cyanide at 50°C and a flowrate of 5 BV/h. The recoveries expected within 15 bed volumes can be seen in Table 5.5 and were subsequently used.

Optimal conditions for the elution of Amberlite PWA 5 loaded with platinum, palladium and gold, with the use of acidic thiourea, were found to be 0.7 M thiourea, 0.2 M HCl at a temperature of 60°C and a flowrate of 5 BV/h.

Base metal elution efficiencies were, however, not determined and due to the lack of any other literature on the elution of base metals from Amberlite PWA 5 data from Fleming & Cromberge (1984*a*, 1984*c*) on the elution of base metals from strong base resins were used. It was assumed to be the same, as Amberlite PWA 5 is also a strong base resin. The recoveries expected within 15 bed volumes can be seen in Table 5.5 and were subsequently used.

Table 5.5: *Expected recoveries of various metals during the elution of ion exchange resins (15 BV)*

Amberlite PWA 5				*Minix*	
Zinc Cyanide		*Acidic Thiourea*		*Acidic Thiourea*	
Metal	*% Recovery*	*Metal*	*% Recovery*	*Metal*	*% Recovery*
Pt	97	Pt	0	Pt	0
Pd	99	Pd	99	Pd	98
Au	65	Au	99	Au	99
Cu	99	Cu	80	Cu	80
Ni	80	Ni	65	Ni	65
Zn	*	Zn	28	Zn	28
Fe	99	Fe	2	Fe	2

* *Not Determined*

Schoeman (2012) and Schoeman *et al*. (2012) reported elution efficiencies for Minix loaded with platinum, palladium and gold, and eluted with acidic thiourea. The expected recoveries within 15 bed volumes can be seen in Table 5.5 and were subsequently used. Base metal recoveries were, however, not determined and it was assumed, based on the recoveries of palladium and gold compared to the elution of Amberlite PWA 5, that the base metal recoveries would be the same as that reported by Fleming & Cromberge (1984*a*, 1984*c*) for the elution of a strong base resin with acidic thiourea. From these studies optimal conditions were found to be 0.7 M thiourea, 0.2 $M\ HCl$ at a temperature of 60°C and a flowrate of 5 BV/h.

All of these studies further mentioned that potassium thiocyanate has the potential to be more effective than zinc cyanide for the elution of all metals from ion exchange resins. However, due to the lack of an industrial viable method for the regeneration of a resin eluted with this medium, it was not considered during the flowsheet development. Research as to the regeneration with a ferrous solution has, however, been shown to be effective in laboratory scale experiments, but has not been employed on an industrial scale (Fleming & Cromberge, 1984*a*; Kotze *et al*., 2005; Marsden & House, 2006).

5.1.3.4 Resin Regeneration

Amberlite PWA 5 eluted with zinc cyanide needs to undergo a regeneration step as the zinc is strongly adsorbed onto the resin, reducing its activity (Fleming & Cromberge, 1984*a*, 1984*c*; Kotze *et al*., 2005; Marsden & House, 2006). This is performed by washing the resin with a 33% H_2SO_4 solution, to remove the zinc, where-after the solution is fed to a caustic bath (containing 20% $NaOH$) to recover the zinc as zinc cyanide for reuse in the next elution cycle. It has been shown that the residual zinc loading can be reduced to 1 000 g/t Fleming & Cromberge (1984*c*).

It was assumed that 90% of the zinc cyanide can be reused, as a bleed stream is required to prevent the build-up of inert species and contaminants. The concentration of zinc cyanide was assumed to be restored by the addition of the stoichiometric requirement of zinc sulphate and sodium cyanide, a method reported to be successful by Fleming & Cromberge (1984*c*).

Resins, both Amberlite PWA 5 and Minix, do not require regeneration if elution is performed with acidic thiourea, as the resin is restored to its original chloride form, as a result of the acidic nature of the eluting agent, which ensures maximum resin activity (Fleming & Cromberge, 1984*a*, 1984*b*, 1984*c*; Kotze *et al*., 2005; Marsden & House, 2006).

5.1.4 Recovery of the base and precious metals from the eluates

5.1.4.1 Thermal Hydrolysis

Recovery of the base and precious metals from the eluates is to be accomplished in an autoclave. The Expected recoveries are given in Table 5.6. Desmond *et al.* (1991) showed that at a temperature of 250°C and a pressure of 40 bar, up to 99.5% of both platinum and palladium can be recovered within an hour. The gold cyanide complex is, however, still very stable at the conditions used and according to Adams *et al.* (1995) only 50% of gold precipitates at about 300°C. Gold only starts to decompose at around 240°C (Adams *et al.,* 1994) and thus a conservative recovery of 20% was selected.

Milhara *et al.* (2007) obtained recoveries for zinc and copper in excess of 80% after 1 hour. These experiments were only conducted at 150°C and 10 bar. It was thus assumed that at the designed conditions (250°C, 40 bar) these recoveries will be in the same range as that observed for platinum and palladium, based on the fact that the relative stability and strength of these base metal cyanide complexes are low and weak, if compared to that of platinum and palladium at 25°C presented in Table 5.1. Nickel and iron were assumed to behave the same as zinc and copper.

Green *et al*. (2004) has also shown that for the more stable thiourea complexes almost complete recovery can be achieved. It was thus assumed that the recoveries during thermal hydrolyses were independent of the nature of the solution, either a cyanide of thiourea medium.

Table 5.6: *Expected recoveries of various metals during thermal hydrolysis (250°C, 40 bar)*

Metal	*Metal Complex*	*% Recovery*
Cu	$Cu(CN)_4^{3-}$	99.5
Ni	$Ni(CN)_4^{2-}$	99.5
Zn	$Zn(CN)_4^{2-}$	99.5
Fe	$Fe(CN)_6^{4-}$	99.5
Pt	$Pd(CN)_4^{2-}$	99.5
Pd	$Pd(CN)_4^{2-}$	99.5
Au	$Au(CN)_2^-$	20.0

5.1.4.2 Electrowinning

For the recovery of the remaining gold by the direct electrowinning of the eluates, a method currently employed in CIP operations, recoveries in excess of 99% were assumed (Fleming, 1992; Stange, 1999;). This value was assumed based on the work performed by van Tonder & Sole (2006). Recoveries in excess of 99% within 5 hours were achieved from various solutions. Electrowinning gold from a thiourea solution, as obtained during typical RIP operations, has also been shown to be effective, at recoveries in excess of 99% within 5 hours (Conradie *et al.*, 1994). An electrowinning residence time of 5 hours was thus used in sizing the equipment.

It was thus assumed that the relative performance of the electrowinning circuits were independent of the nature of the eluate, either deionised water, zinc cyanide or acidic thiourea. This was based on the aforementioned literature and the fact that the destruction of the thiourea and cyanide complexes during thermal hydrolysis is believed to ease the operation of electrowinning.

Due to the very small amounts of the other base and precious metals still remaining in the solutions coming from the autoclaves (only about 4% than of the gold content), it was assumed that the majority (98%) of these metals will be co-deposited along with the gold. No published data were found on the electrowinning of platinum and palladium, thus whether these metals could be recovered using this technique are still unknown at present.

Chapter 6

Process Economics

The comparison between the different flowsheet alternatives will be primarily based on the process economics. As this project was on a concept phase level, a desktop study economic analysis was followed, which typically corresponds to an error range of ±30% (Peters & Timmerhaus, 1991). This included the preliminary design and sizing of the process equipment to be used, capital and operating cost estimations for all flowsheets and net present value calculations used for flowsheet comparison. All estimations were based on designing a process for the treatment of the leach solutions specified in Table 4.1, thus a combined pregnant leach solution feed of 354 t/h which corresponds to a mining operation of 240 000 t/m.

This chapter focuses on the methods used to determine these different economic parameters. Subsequent comparisons between the different flowsheet alternatives based on the costs estimations are discussed in Chapter 8.

6.1 Capital Cost Estimations

6.1.1 Overview

As this study was only at a concept/feasibility phase, an order of magnitude approach was followed to estimate the capital costs. The major process equipment cost ratio method, mentioned by several authors (Peters & Timmerhaus, 1991; Smith, 2005; Turton *et al.*, 2010) was used. This method uses cost factors, based on the delivered cost of the major process equipment, to estimate the direct and indirect capital costs of a project.

From the process flow diagrams the major process units were identified and sized with the use of the mass balance results. Preliminary costs where then obtained, either by estimating with the use of empirical cost correlations or supplier quotes. These costs were then used, together with the cost factors, to estimate the total capital expenditure of all the flowsheet alternatives. These methods are discussed in the following sections.

6.1.2 Equipment Design and Sizing

This section focuses on the equipment design and sizing decisions that were made. For detailed equipment sizes, materials of construction and the designed operating conditions, please refer to Appendix B.

6.1.2.1 SART

Total plant costs for the construction of a typical SART plant was used in the estimation of the SART plant costs. Stewart & Kappes (2012) stated that a typical SART plant can cost anywhere from \$750 - \$1 500 per cubic meter of solution treated per day (m^3/d). An average cost of \$1 200 per m^3/d was assumed for the SART plant treating the combined leach solution and was scaled to present day value with the use of cost indexes, while the costs of the separate PLS SART plants were scaled with a scaling factor of 0.6. This scaling procedure is discussed in section 6.1.3. As the total solution throughput of the SART plants are the same for all flowsheet alternatives, the assumed value would have no significant influence on the comparison between the different flowsheet alternatives. Based on this, these costs were assumed to be acceptable.

6.1.2.2 Adsorption Sections

The adsorption circuits were based on the AAC Pump Cells mode of operation. This was due to the significant cost and operating benefits thereof as opposed to conventional CIP tanks (Stange, 1991; Stange, 1999; Rogans & McArthur, 2002). The equipment cost of such an adsorption circuit was obtained from the sole supplier of these process units, Kemix PTY Ltd. The cost of each of the adsorption sections were then scaled according to the size required.

The adsorption sections were sized according to the type of adsorbent used. For the carbon adsorption sections a solution residence time of 20 minutes per stage were selected. This is the typical residence time used in carbon adsorption pump cell operations (Stange, 1991; Rogans & McArthur, 2002; Fleming *et al.*, 2011).

Fleming (1993) mentioned that resin adsorption tanks could be up to 5 times smaller than that required in activated carbon adsorption circuits, due to the higher mass to volume ratio of resins and the superior gold adsorption kinetics. This was, however, based on conventional adsorption circuits, and it was believed that the difference in tank sizes would be less pronounced in pump cell circuits, as these already operate at very short residence times.

It was thus assumed that for the resin adsorption sections, a solution residence time of 10 minutes would be adequate. This was based on the fact that the kinetics of adsorption for resins is approximately twice those of carbon.

Tank sizes were subsequently calculated based on these residence times and solution throughput in each specific section.

6.1.2.3 Elution Sections

The elution sections were sized based on the assumption that one elution cycle is performed per day. The acid wash and elution columns were thus sized according to the amount of carbon transferred per day for each of the individual flowsheet alternatives. Raw material feed tanks, such as those for fresh eluent and acid, were sized to hold the equivalent amount required for one elution cycle to be performed.

The material of construction for the acid wash and elution columns were selected to be Hastelloy C as this material can withstand conditions of high temperature and pressure and exhibit excellent corrosive resistance against hydrochloric acid and sulphuric acid (Marsden & House, 2006; Turton *et al.*, 2010).

Acid storage and feed tanks were also selected to be manufactured from Hastelloy C while the pre-treatment and eluate tanks were selected to be manufactured from the more cost effective stainless steel 316, as these units do not operate under such extreme conditions.

6.1.2.4 Recovery Sections

The subsequent recovery sections were sized based on treating one batch of eluate per day, as only one elution cycle is to be performed per day as mentioned above.

The eluate hold-up tanks prior to final metal recovery were sized to hold a full batch of eluate, while the hold-up tank between thermal hydrolysis and electrowinning were sized to hold one thermal hydrolysis batch. Stainless steel 316 was selected as the material of construction.

Thermal hydrolysis is a batch process and the autoclaves were sized to treat the maximum number of batches possible within one elution cycle. This was done to ensure that relatively small autoclaves were used, as the equipment cost thereof was significant. One batch required 1 hour, as discussed in Chapter 5, while a further 20 minutes were allowed to account for filling and draining of solution and precipitate.

Costs for an electrowinning circuit were obtained from the supplier, Kemix PTY Ltd, while flowsheet specific costs were scaled according to solution throughput. Electrowinning was assumed to be a continuous process with an average solution residence time of 5 hours required to achieve the desired extraction. This allotted time has been shown to be sufficient in achieving extractions in excess of 99% for both carbon eluates (van Tonder & Sole, 2006) and resin eluates (Conradie *et al*., 1994).

6.1.2.5 Regeneration

It was assumed that carbon regeneration would be performed in a rotary kiln regeneration circuit. Costs for such circuits were obtained from Zanitsch & Lynch (1978) and were used in estimating the costs for each individual section based on scaling with regards to the carbon throughput to the reactivation circuit per day. The steam and fuel requirements were assumed to be available on site, and were included in the overall utility cost factors.

Regeneration of the Amberlite PWA 5 resin, eluted with zinc cyanide, was assumed to be performed in the elution columns, to allow for cost savings. Thus the only equipment required were the feed tanks of the chemicals used during regeneration. Hastelloy C was selected as the material of construction for the acid tanks while stainless steel 316 was selected for the neutralization tanks. These tanks were sized to hold the equivalent amount required for one regeneration cycle.

6.1.3 Estimation of the Purchased Equipment Cost

Costs for the specific process equipment, based on the design and size thereof, were either obtained from suppliers or estimated with the use of empirical correlations, such as those provided in Turton *et al.,* (2010). Detailed costs can be found in Appendix C.

6.1.3.1 Estimated Equipment Cost Correction

The estimated purchased equipment costs obtained by the use of empirical correlations were, however, at base case conditions (C_P^o), i.e. constructed with carbon steel and capable of withstanding only ambient temperatures and atmospheric pressure. Correction factors were applied in order to account for the extra cost of the more expensive materials of construction used and the costs required to ensure that the equipment can withstand the more rigorous operating conditions.

These correction factors are given in Table 6.1 and were applied according to equation (6.1) (Smith, 2005; Turton *et al*., 2010).

Table 6.1: *Correction factors used in equipment cost estimations (Smith, 2005; Turton et al., 2010)*

Material of Construction		*Temperature*		*Pressure*	
Material	F_M	$T\ (°C)$	F_T	$P\ (Bar\ abs)$	F_P
Carbon Steel	1.0	0 - 100	1.0	0.1 - 0.5	1.3
Stainless Steel 316	1.8	100 - 300	1.6	0.5 -7	1.0
Nickel (Monel)	3.6	300 - 500	2.1	7 - 50	1.5
Hastelloy C	5.8			50 - 100	1.9

$$Purchased\ equipment\ cost = C_P^o \times F_M \times F_T \times F_P \quad (6.1)$$

where

F_M = Correction factor for material of construction

F_T = Correction factor for temperature

F_P = Correction factor for pressure

6.1.3.2 Estimated Equipment Cost Scaling

If costs were available at a different size of the respective unit, equation (6.2) was used to scale it to the correct size. Scaling factors were obtained from Smith (2005) and Turton *et al.* (2010) and can be found in Appendix C.

$$C_A = C_B \times \left(\frac{S_A}{S_B}\right)^n \quad (6.2)$$

where

C_A = Estimated cost of equipment at size A

C_B = Known cost of equipment at size B

S_A = Equipment size A

S_B = Equipment size B

n = Scaling Factor

If costs were available from another year, equation (6.3) was used to scale to the present day cost. The Chemical Engineering Plant Cost Index (CEPCI) was used for scaling of these costs.

$$C_2 = C_1 \times \frac{I_2}{I_1} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (6.3)$$

where

C_1 = Known cost of equipment at year 1

C_2 = Estimated cost of equipment at year 2

I_1 = Cost Index at year 1

I_2 = Cost Index at year 2

6.1.4 Fixed Capital Investment

As mentioned earlier, the major process equipment cost ratio method was used in estimation of the total fixed capital costs of the different flowsheet alternatives. Table 6.2 shows the capital cost breakdown and the respective cost factors for each individual cost used in this project. The cost factors presented in Table 6.2 are based on a processing plant treating a combination of solids and fluids. These cost factors were used as the proposed flowsheet alternatives all entail solid processing steps and fluid processing steps. A 3% delivery charge was added to the purchase equipment costs, as estimated based on the methods discussed in the previous section, to estimate the delivered equipment costs (Peters & Timmerhaus, 1991). These cost factors were applied to the delivered equipment cost to ultimately arrive at an estimate of the total capital cost, or in other words the fixed capital investment.

Table 6.2: *Capital cost factors based on the delivered equipment cost (Peters & Timmerhaus, 1991)*

Cost Component	*% Of Delivered Equipment Cost*
Direct Costs	
Delivered Equipment Cost	1.00
Equipment Installation	0.39
Piping (Installed)	0.31
Instrumentation & Controls	0.13
Electrical (Installed)	0.10
Utilities	0.35
Off-sites	0.20
Buildings (including services)	0.29
Site Preparation	0.06
Total Direct Cost	***2.83***
Indirect Costs	
Design, Engineering and Supervision	0.32
Construction Expenses	0.34
Total Indirect Cost	***0.66***
Contractors Fee	5% †
Contingency	25% †
Total fixed capital cost	***4.54***
Working Capital	15% ‡
Total Capital Cost	***5.34***

† *% of Total fixed capital cost*

‡ *% of Total capital cost*

Detailed capital cost estimates for each flowsheet alternative can be found in Appendix C.

6.2 Operating Cost Estimations

6.2.1 Overview

Operating costs for the different process flowsheets were estimated based on a ratio method used in the chemical/process engineering industry (Turton *et al.,* (2010). In this method, cost factors based on estimates of the capital, raw materials, utilities and labour costs are used to arrive at an estimate of the overall operating costs. These costs are estimated beforehand after which the cost factors are applied. A breakdown of the operating costs, and the respective cost factors are presented in Table 6.3.

Table 6.3: *Operating cost factors (Turton et al., 2010)*

Cost Component	*Cost Factor*
Direct Operating Costs	
Raw Materials (C_{RM})	1 C_{RM}
Waste Treatment (C_{WT})	1 C_{WT}
Utilities (C_{UT})	1 C_{UT} 0.2 C_{TOC}
Operating Labour (C_{OL})	1 C_{OL}
Direct Supervisory and Labour	0.18 C_{OL}
Maintenance and Repairs	0.06 FCI
Operating Supplies	0.009 FCI
Laboratory Charges	0.15 C_{OL}
Patents and Royalties	0.03 C_{TOC}
Fixed Operating Costs	
Depreciation	†
Local Taxes and Insurance	0.032 FCI
Plant Overhead Costs	0.708 C_{OL} 0.036 FCI
General Operating Expenses	
Administration Costs	0.177 C_{OL} 0.009 FCI
Distribution and Selling Costs	0.11 C_{TOC}
Research and Development	0.05 C_{TOC}
Total Operating Cost (C_{TOC})	*1* C_{TOC}

† *Depreciation was handled separately*

FCI *Fixed Capital Investment*

A contingency of 25% was added to allow for unforeseen expenses. Detailed operating cost estimates for each flowsheet alternative can be found in Appendix D.

6.2.2 Estimation of Raw Material Costs

Estimation of the raw material costs were performed by multiplying the required amounts, as obtained from the mass balances, with the cost of each of the respective chemicals used. Costs were obtained from either literature or suppliers. A 20% levy was added to account for delivery costs of the raw materials. These costs are presented in Table 6.4. Detailed raw material costs for each flowsheet alternative can be found in Appendix D.

Table 6.4: Raw material costs

Raw Material	*Formula*	*Purity (%)*	*Cost (US$/t)*
Sulphuric Acid	H_2SO_4	98	360†
Sodium Hydrosulphide	$NaHS$	70	600†
Lime	$Ca(OH)_2$	93	132†
Flocculant (Hychem AF 305)	-	100	1 800†
Hydrochloric Acid	HCL	33	120†
Sodium Cyanide	$NaCN$	99	660†
Caustic Soda	$NaOH$	98	1 320†
Zinc Sulphate	$ZnSO_4$	100	840†
Thiourea	CH_4N_2S	100	1 980†
Process Water	H_2O	100	1.47‡
Deionised Water	H_2O	100	4.42‡
Carbon	-	-	1 920†
Resin	-	-	18 829¥

† *Alibaba (2013)*
‡ *Turton et al. (2010)*
¥ *van Deventer et al. (2000)*

6.2.3 Estimation of Waste Treatment Costs

Waste treatment costs were estimated by characterizing the respective waste streams into three groups, either no-treatment required, non-hazardous or hazardous. The treatment costs per ton for each category, obtained from Turton *et al.* (2010) and scaled to present day values, are given in Table 6.5.

Table 6.5: Waste treatment costs

Treatment Required	*Cost (US$/t)*
No Treatment	0
Non-hazardous	41
Hazardous	104

Treatment costs for each of the waste streams were assigned and combined with the respective mass balance results to obtain the waste treatment costs of each flowsheet alternative. The category and costs of the waste streams can be found in Appendix D.

6.2.4 Estimation of Utility Costs

Utility costs were estimated by combining the mass balance results with energy balances. It was assumed that boiler efficiencies were 80% (CIBO, 1997) and that high pressure steam (254°C, 41 barg) was available for use (Turton *et al.*, 2010). For regeneration of the carbon in the rotary kiln it was assumed that fuel will be used to fire the kiln and that regeneration occurs in a steam environment to minimise losses due to oxidation. Utility requirement of the regeneration circuits were obtained from Zanitsch & Lynch (1978). It was further assumed that all reagent feed streams were available at 25°C. Power requirements were estimated at 20% of the total operating cost. Utility costs, obtained from Turton *et al.* (2010) and scaled to present day values, are given in Table 6.6. Detailed utility requirements for each flowsheet alternative can be found in Appendix D.

Table 6.6: Utility costs

Utility	*Cost*	
Fuel	16.32	$/GJ
Steam	34.45	$/t
Electricity	0.092	$/kWh
Cooling Water	17.01	$/kg

6.2.5 Estimation of Operating Labour Costs

The number of operators per shift required was estimated using the equation below (Turton *et al.*, 2010).

$$N_{OL} = \left(6.29 + 31.7P^{0.1} + 0.23N_{np}\right)^{0.5} \quad (6.4)$$

where

N_{OL} = The number of operators required per shift

P = The amount of particulate processing steps

N_{np} = The amount of non-particulate processing steps

The particulate processing steps for each plant section involves the physical handling or transportation of solid material. These were selected to be the base metal filter and gypsum filter operation during SART, the removal of the precious metal precipitate during thermal hydrolysis and the removal of the gold precipitate during electrowinning.

Non-particulate processing steps involve the automated flow or solution/pulp or solids during plant operation. These were selected based on the major processing steps involved in each plant section and are listed in Table 6.7. Although the adsorption section consists of multiple tanks, it was assumed that the operation thereof will be similar and thus a reduced number were used.

Table 6.7: *Major process units per plant section*

Plant Section	*Processing Steps*	*Units*
SART	4	Reactors and thickeners
Adsorption	2	Adsorption tanks
Elution	2	Acid wash and Elution columns
Thermal Hydrolysis	1	Autoclave
Electrowinning	1	Electrowinning cells
Regeneration	1	Rotary kiln

Turton *et al.* (2010) has stated that for each operator required per shift, 4.5 operators should be hired for a typical 8 hour shift and 49 week operating cycle per year. Rogans & McArthur (2002) reported operator salaries for a typical South African mine which was scaled, at 6% per year, to present day values. This amounted to an operator salary of US$ 10 700 per annum.

The operating labour was thus estimated using:

$$C_{OL} = 4.5\, N_{OL} \times \$\,10\,700 \quad (6.5)$$

6.3 Project Profitability

The profitability of the different flowsheet alternatives are to be compared with the use of the net present value (NPV). The NPV is the cumulative cash flow at the end of the entire project life and gives an indication as to what the financial prospect of a project could be, either favourable or it could result in major losses. The different financial parameters needed in order to estimate the NPV are discussed below.

6.3.1 Fixed Expenses

The fixed capital investment, as discussed in section 6.1, forms part of the initial costs required to start-up the project. This together with the working capital, required for initial salaries and operating expenses, contribute to the overall fixed expenses at the start of the project.

Another very important parameter used in carbon and resin adsorption circuit evaluation, that affects the initial cash flows of the project, is the value mineral lock-up. The value mineral lock-up is the amount of valuable material, such as the platinum, palladium and gold, constantly loaded on the carbon or resin particles that are within the adsorption circuit. This material loads onto the carbon or resin in the initial stages of adsorption and is locked up within the plant and is only recovered at project decommissioning. It is thus imperative that this value is minimised.

The value mineral lock-up plays a significant role during plant design and operation. The trade-off between the initial revenue losses and the overall capital and operating costs gained for operating at a specific lock-up need to be carefully taken into consideration when operating conditions are decided.

The total fixed expenses of the project are thus the cost of land, capital cost, working capital, carbon or resin inventory costs (first fill) and the allowance for value mineral lock-up.

6.3.2 Variable Expenses and Income

The operating cost, as discussed in section 6.2, is the major expense paid every year throughout the project life. This together with the depreciation and revenue form part of the variable expenses and income each year and directly influences the cash flow.

6.3.2.1 Depreciation

Depreciation of capital was calculated using the straight line method over an assumed equipment life of 8 years. The straight line method used is given by equation by (6.6).

$$d_k^{SL} = \frac{[FCI_L - S]}{n} \quad \ldots\ldots\ldots\

6.3.2.2 Potential Project Income

The proposed flowsheets are part of a bigger overall process, thus direct revenues are not obtained, as products that still need to undergo refining are produced. A potential income based on the contained metal value of each of the product streams was therefore used in assessing the possible income that could be obtained by the different flowsheet alternatives.

For the base metal precipitates, produced during the SART process, a value of 50% of the contained metal value was assumed. This was due to the fact that the metals are still in a sulphide form and extensive treatment would still be required to recover the metals in their metallic form. It is furthermore a mixed product consisting of various metals. For the gypsum precipitate, also produced during SART, an assumed value of 100% was used as it was an almost pure gypsum product that could be sold directly.

For the precious metal precipitates, produced during thermal hydrolysis, a value of 80% of the contained metal value was assumed. This is because the precipitates were of high grade (>80%) and that the metals are assumed to be recovered in their respective metallic state. Further refining of these products would, however, still be required. The same assumption was made for the gold electrowinning product.

The metal prices used in calculating the potential income of the different process alternatives are presented in Table 6.8. In order to account for fluctuations in these prices and to ensure that it is not biased, the average market prices over the 6 month period May 1st - October 31st 2013 were used (InfoMine, 2013).

Table 6.8: Metal Prices

Metal	Price (US$/kg)
Cu	7.23
Ni	13.78
Zn	1.85
Fe	0.13
Pt	45 610
Pd	23 450
Au	43 000

6.3.3 Net Present Value

The Net Present Value (NPV) has been calculated for each process based on a 15 year project life. It should be noted that the calculated NPVs provide a basis for comparison but do not represent the complete cash flow at the end of the projects, as the costs of mining, leaching and final metal refining were not included. It was assumed that these additional costs of mining, leaching and refining would be the same for all of the different flowsheet alternatives and were thus omitted from the actual calculations. The calculated NPVs thus only served as a comparative measure used in comparing the financial prospects of the different flowsheet alternatives to one another.

The assumptions made during the calculation of the NPVs are listed below:

- Cost of land was assumed to be negligible, as it would be the same for all process alternatives.
- Construction of the processing plants was assumed to be completed after 2 years and that the working capital and value mineral lock-up penalty is paid at the start of year 3.
- All cash flows were discounted at a typical rate of 15%.
- It was assumed that the plants would take up to 2 years to achieve full capacity and that during start-up an average of 50% and 80% of the potential revenue were reached after years 3 and 4 respectively (operating year 1 and 2).
- A fixed taxation rate of 45% was assumed.
- A salvage value of 10% of the initial plant costs were assumed and could be recovered at the end of the project, together with the initial working capital.
- Value material locked up within the plant were also summed to be recovered at the end of the project and sold at market value.
- The plants were assumed to be shut down for a period of 4 weeks each year to allow for maintenance. The plants thus had an 8088 hour operating year.

Detailed discussions on the different flowsheet alternatives and their respective NPVs are given in Chapter 8.

Chapter 7

Adsorption Modelling and Optimization

It is well known that one of the most important sections of a CIS or RIS operation is the adsorption circuit (Stange, 1991, 1999). Its contribution to the overall capital and operating costs are significant and its performance directly influences the size of the downstream elution and regeneration circuits and their performance. Furthermore any gold or value material not extracted during the adsorption process is lost to the tailings. An optimum design of the adsorption circuit is thus required to ensure that such a plant is operated at its full potential, both from a metallurgical and economic perspective. Kinetic models can be used to simulate different scenarios or adsorption circuits and could be used to size these circuits. An optimal design can then be achieved by combining such a model with economic models. In this chapter a review of existing kinetic models for gold adsorption are presented followed by a typical method of optimization.

7.1 Adsorption Modelling Review

Kinetic models for modelling of the carbon-in-pulp (CIP) process have been reported by several researcher since the late 1970's (see e.g. Dixon *et al.*, 1978; Fleming *et al.*, 1980; Nicol *et al.*, 1984*a*, 1984*b*; Johns, 1986; Stange & King, 1987; Woollacott *et al.*, 1990; Liebenberg & van Deventer, 1998). The following section is a review of three of the most widely used adsorption models for the CIP process.

7.1.1 Nicol-Fleming Model (1984)

One of the first, and simplest models, is the one developed by Nicol *et al.* (1984*a*, 1984*b*). The model assumes that all tanks can be modelled as perfectly mixed reactors and that a steady state condition with counter-current flow of both carbon and solution is achieved, such that the gold concentration in the solution and that on carbon can be assumed to be constant within each adsorption tank.

The model was extended to allow for periodic transfer of carbon while a kinetic leaching model was incorporated to allow for the simulation of carbon-in-leach (CIL) operations. These models were found to yield reasonable results when compared to plant data.

7.1.2 Stange Model (1990)

This model was first presented in a paper by Stange & King (1987) and was later presented in more detail in a series of three papers (Woollacott *et al.*, 1990; Stange *et al.,* 1990*a*, 1990*b*).

A population-balance approach was used along with a linear kinetic adsorption rate from the Dixon model (Dixon *et al.*, 1987). It takes into account the gold loading of individual particles in each adsorption stage allowing for the distribution of adsorption rates and gold loadings on the carbon particles to be determined. Non-instantaneous transfer of carbon is incorporated along with carbon leakage from tank to tank. This was an improvement, as previous models (Dixon *et al*., 1978; Nicol *et al.*, 1984*b*) assumed instantaneous transfer of carbon without back mixing.

A model similar to the Stange model was produced by Carrier *et al.* (1987). This model does not take into account the distribution of gold loadings and uses differential mass-balances incorporated with the Dixon adsorption model.

The Stange model was validated through testing and three different case studies, using different scenarios to assess the performance of the model. It was found that the model was in good accordance to actual plant data and could be used for a wide range of operating conditions (Staunton, 2005).

7.1.3 Liebenberg & van Deventer Model (1998)

A CIP model was developed by Liebenberg and van Deventer (1998) that incorporates film and intra-particle diffusion, simultaneous leaching and loading of multicomponent species together with carbon fouling and changes in the cyanide through the adsorption circuit and as well as changes in the loading isotherms between tanks. It is based on the branched pore model in which adsorption occurs by diffusion through a liquid film surrounding a carbon particle and diffusion inside the carbon pores from the macro-pores to the micro-pores.

7.1.4 Model Selection

In order to ensure that the solution of the adsorption model was tractable within the context of a plant-wide optimisation study, the relatively simple approach of the Nicol-Fleming model was selected in order to model the adsorption circuits. Although this model is not the most accurate and some bold assumptions were made in its development, it still predicts the gold in solution and gold on carbon profiles very well, as will be discussed in the next sections. The other models might yield more accurate results but as this study formed part of the initial process development, based on experimental tests that were primarily designed to establish the feasibility of these methods, the experimental data were insufficient for the more complex and rigorous models to be used. Furthermore this model can be combined with economic models with relative ease to arrive at the optimum operating conditions.

Previous research (Mpinga, 2012) has, however, suggested that a pseudo 2nd order model be used for modelling the adsorption of platinum and palladium onto activated carbon. Such a model was available from literature (Ho & McKay, 1999) and a 2nd order model analogous to the Nicol-Fleming model was developed. There were, however, difficulties in implementing these models within the simulator, partly due to the complexity thereof and the limited data available. A discussion on the work performed regarding these models is given in Appendix E.

7.1.5 Nicol-Fleming Model Derivations

This section is a review of the three paper series published by Nicol *et al.* (1984*a*, 1984*b*, 1984*c*) on the development of their CIP adsorption model and its application.

7.1.5.1 Model Background

The model is based on consideration of the boundary between the pregnant solution and the activated carbon as an interface across which the gold, in the form of the aurocyanide ion ($Au(CN)_2^-$), is transferred. It is assumed that equilibrium is achieved at this interface and that the mass transport of the gold into the carbon particle can be characterised by a single pseudo-mass-transport coefficient (k_c) and for the solution phase a corresponding coefficient (k_s). A further assumption is made that the equilibrium between the gold in solution and that on the carbon surface can be described by equation (7.1):

$$[Au]_c^s = K[Au]_s^s \quad \ldots\ldots\ldots\

where

$[Au]_s^s$ = The concentration of gold in solution at the carbon surface (g gold/t solution)

$[Au]_c^s$ = The gold on carbon loading at the carbon surface (g gold /t carbon)

K = The Equilibrium Constant (t solution/ton carbon)

This assumption is reasonable as most adsorption circuits exhibit linear isotherms at low gold concentrations. Assuming that the concentration gradients in both phases are linear, the rates of change of gold concentration in each phase can be written as:

$$\frac{d[Au]_s}{dt} = \frac{-k_s A M_c}{M_s}(K[Au]_s - [Au]_s^s) \quad . . . . . (7.2)$$

$$\frac{d[Au]_c}{dt} = k_c A([Au]_c^s - [Au]_c) \quad . . . . . (7.3)$$

where

$[Au]_s$ = The concentration of gold in solution (g gold/t solution)

$[Au]_c$ = The gold on carbon loading (g gold/t carbon)

A = The superficial area per unit volume of carbon (m^2/m^3)

M_c = The mass of carbon (t)

M_s = The mass of solution (t)

t = Time (h)

Equating the rates of transfer and substitution of equation (7.1) yield the following rate expressions:

$$\frac{d[Au]_s}{dt} = \frac{-kM_c}{M_s}(K[Au]_s - [Au]_c) \quad . . . . . (7.4)$$

$$\frac{d[Au]_c}{dt} = k(K[Au]_s - [Au]_c) \quad . . . . . (7.5)$$

where

$$k = \frac{k_s k_c A}{k_c K + k_s}$$

These rate expressions can be combined with mass balances to enable the derivation of models that can be used to predict either batch operations or continuous operations. The derivation of these two models is discussed next.

7.1.5.2 Batch Adsorption Kinetic Expression

In a batch adsorption test a known quantity of activated carbon is added to a stirred contactor containing a known quantity of a gold rich solution. A mass balance over such a system is given by equation (7.6).

$$[Au]_c = [Au]_{c,o} + \frac{M_s}{M_c}\left([Au]_{s,o} - [Au]_s\right) \quad . . . . . (7.6)$$

where

$[Au]_{s,o}$ = The concentration of gold in solution at time zero (g gold/t solution)

$[Au]_{c,o}$ = The gold on carbon loading at time zero (g gold/t carbon)

Substitution of this mass balance into the rate equation, equation (7.4), and integrating at the boundary conditions $t = 0$ to $t = t$ and $[Au]_s = [Au]_{s,o}$ to $[Au]_s = [Au]_s$ and noting that for a batch system $[Au]_{c,o} = 0$, the model reduces to:

$$\frac{ln\left([Au]_s KM_c + [Au]_s M_s - [Au]_s M_{s,o}\right) - ln\left(KM_c[Au]_{s,o}\right)}{(KM_c + M_s)} = \frac{-k}{M_s}t \quad . . . . . (7.7)$$

Rearrangement and simplification of equation (7.7) yields the batch adsorption kinetic model, given by equation (7.8):

$$ln\left(\frac{[Au]_s - \beta}{[Au]_{s,o} - \beta}\right) = -k\alpha t \quad . . . . . (7.8)$$

where

$$\alpha = \frac{KM_c}{M_s} + 1 \quad ; \quad \beta = \frac{M_s[Au]_{s,o}}{(KM_c + M_S)}$$

Figure 7.1 shows various batch adsorption test data, carried out on pulp from the Grootvlei Gold Mine, with equation (7.8) fitted to the data (Nicol *et al.*, 1984*a*). It can be seen that the model accurately predicts the gold in solution concentration for different conditions. It should be noted that the kinetics of the rate of gold adsorption are dependent on the mixing efficiency (Nicol *et al.*, 1984*c*) as can be seen from the decrease in performance when using larger air agitated tanks as opposed to the smaller mechanically stirred batch reactors.

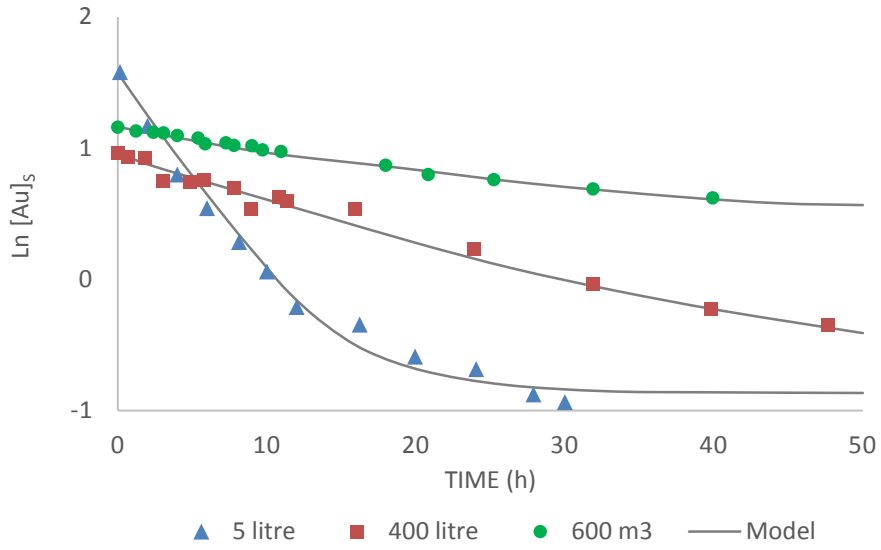

Figure 7.1: *Batch adsorption test data carried out on pulp from the Grootvlei Gold Mine fitted with equation 7.8. (Redrawn from Nicol et al., 1984a)*

From a non-linear least squares fit of equation (7.8) to batch adsorption test data, such as that presented in Figure 7.1, the rate constant (k) and equilibrium constant (K) can be determined that are required for use in modelling a counter-current CIP adsorption circuit, as explained in the next section.

7.1.5.3 Counter-Current Steady State Adsorption Model

This model makes the assumption that the flow of pulp/solution and carbon is counter-current and continuous, and that steady state conditions are reached in which the concentration of gold in solution and on the activated carbon is constant in each stage of the adsorption circuit. A further assumption is made that each stage can be treated as a perfectly mixed stirred reactor with constant residence times for both the pulp/solution and carbon.

As discussed in the previous section the rate of adsorption of gold onto activated carbon can be described by:

$$\frac{d[Au]_c}{dt} = k(K[Au]_s - [Au]_c) \quad . . . . . \quad (7.5)$$

This rate equation can be combined with mass balances over an entire counter-current CIP adsorption circuit to yield a model that can be used to describe the performance of such a circuit. Figure 7.2 is a schematic representation of a typical counter-current CIP adsorption circuit consisting of N stages.

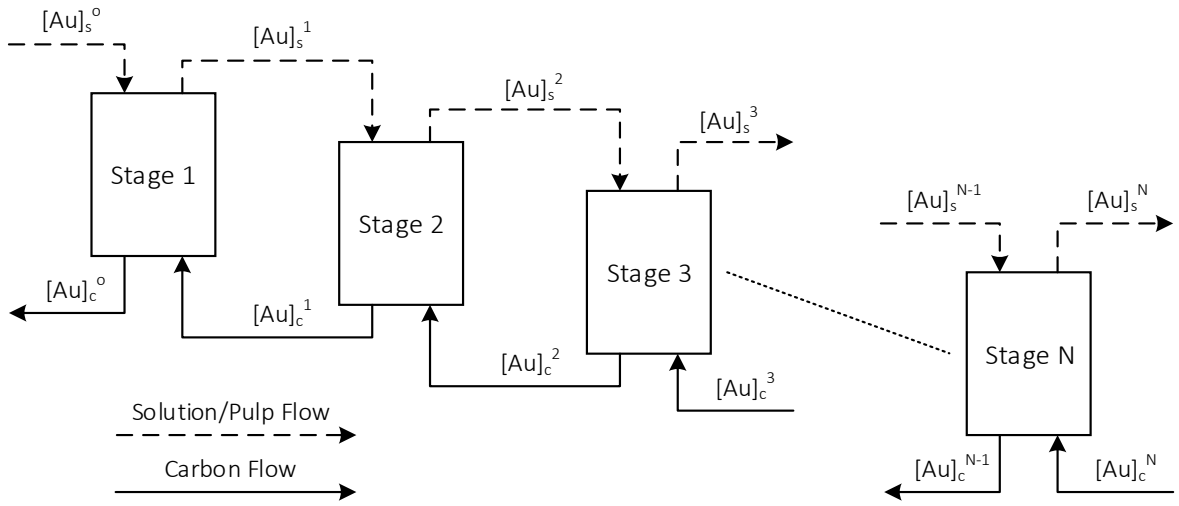

Figure 7.2: *Schematic representation of a typical counter-current CIP adsorption circuit*

The overall mass balance for gold across the entire circuit is represented by:

$$V_s([Au]_s^o - [Au]_s^N) = V_c([Au]_c^o - [Au]_c^N) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (7.9)$$

where

V_s = The flowrate of the feed solution (m^3)

$[Au]_s^o$ = The gold concentration in the feed solution (g gold/t solution)

$[Au]_s^N$ = The gold concentration in solution leaving stage N (g gold/t solution)

V_c = The flowrate of activated carbon (m^3)

$[Au]_c^N$ = The gold loading on carbon entering stage N (g gold/t carbon)

$[Au]_c^o$ = The gold loading on carbon leaving the first stage (g gold/t carbon)

Consideration of this mass balance across the first stage and the rate equation occurring in this stage, the following equation can be used to predict the metallurgical performance:

$$[Au]_c^o = \left(\frac{[Au]_c^1 + K[Au]_s^1 kt_c}{1 + kt_c}\right) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (7.10)$$

where

t_c = The residence time of carbon per stage (h)

Using this equation an iterative approach can be used to determine the residence time of the carbon per stage, as this is the only variable unknown if a specific metallurgical performance, i.e. gold on carbon loading and barren solution concentration, is to be achieved. This iterative approach, as presented by Nicol *et al.* (1984*b*), is discussed below.

A first estimate, say 5 hours, is made for t_c, which is used in the calculation of $[Au]_s^1$ and $[Au]_c^1$ from the simultaneous solution of equation (7.10) and the mass balance across the first stage. These values can then be used in the calculation of the corresponding quantities for the second stage, and so on until the values of $[Au]_s^N$ and $[Au]_c^N$ are obtained. If the value for $[Au]_s^N$ is not as desired, a new estimate of t_c is made. This approach is repeated until $[Au]_s^N$ converges to the value required as to meet the selected metallurgical performance.

With a pre-determined solution residence time, usually an hour per stage for conventional CIP operations, the size of the adsorption tanks is known and the residence time of the carbon (t_c) can be used in determining the mass of carbon per stage. This allows for the calculation of the carbon concentration per stage, the total carbon inventory and its associated gold lockup.

7.1.5.4 Model Performance

Figure 7.3 shows a comparison between actual pilot plant data and values estimated by the above mentioned model for a batch adsorption test conducted at the Western Areas Gold Mine (Nicol *et al.*, 1984*b*). It can be seen that the gold solution profile estimated by the model were very similar to that observed in the pilot plant.

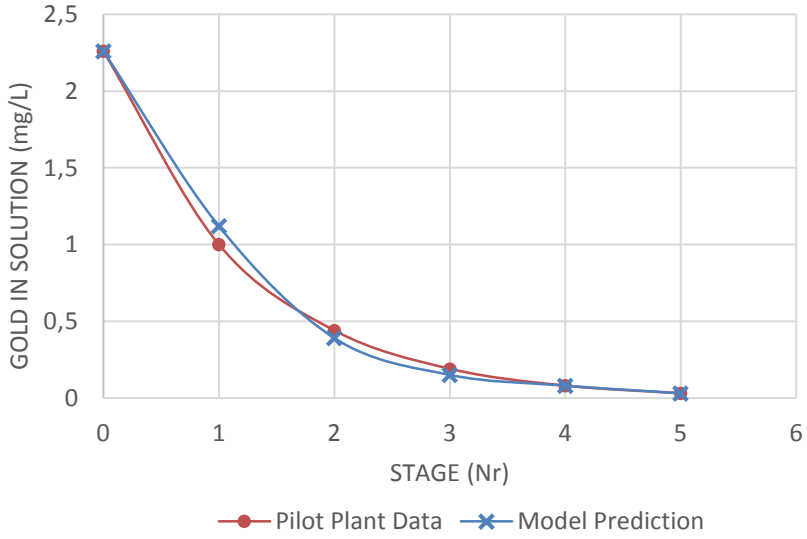

Figure 7.3: *Comparison between pilot plant data and model predictions (After Nicol et al., 1984b)*

The amount of carbon per stage necessary to obtain the required metallurgical performance was, however, under predicted. The observed value was about 10 kg per stage while the model estimated 6.4 kg. This was attributed to the fact that gold from the ore is dissolved and desorbed during CIP adsorption, factors not taken into account by the model, and that the model is very sensitive to changes in the barren solution concentration. A decrease in the barren concentration from about 0.03 to 0.02 g/t resulted in a prediction of 8.3 kg per stage.

Carbon adsorption circuits, however, seldom operate in a true continuous flow of carbon. A fraction of the carbon inventory per stage is rather transferred from one adsorption tank to the next at a specified time, i.e. there is a periodic transfer of carbon. A model that incorporates this mode of operation was also developed by Nicol *et al.* (1984*b*). It is, however, more complex, as the steady state assumption made in the continuous flow model no longer holds. The model is thus a complex set of differential equations, increasing the difficulty in utilising the model.

Table 7.1 shows the relative performance of an adsorption circuit operating under different carbon transfer conditions, as calculated by the Nicol-Fleming Models, and serves as a comparison between continuous transfer and the periodic transfer of carbon. It can be seen that there is very little difference in the estimated performance between the different modes of operation.

Table 7.1: *The effect of different carbon transfer modes of operation on the relative performance of a counter-current CIP adsorption circuit (After Nicol et al., 1984b)*

Fraction transferred per cycle	*Relative Performance (%)*
1.00	100
0.50	95
0.25	92
0.10	90
Continuous	88

The continuous model was selected for use in this project, as its ease of application was a major benefit. However, it under predicts the performance that could be expected, as the AAC Pump Cell mode of operation is proposed for the adsorption sections. The AAC Pump Cells work on the principle of complete periodic transfer of the carbon by changing the solution feed stage as opposed to the physical transfer of the carbon from stage to stage, i.e. the carousel mode of operation (Whyte *et al.*, 1990; Rogans & McArthur, 2002; Dippenaar & Proudfoot, 2005).

As this is a feasibility/concept phase study and the economic analysis is performed at an order of magnitude scale, the error of utilizing this model was deemed insignificant. It has furthermore been shown (Williams & Glasser, 1985) that the relative performance between a CIP and a CIS circuit is very similar. The model was thus used in the simulation of the CIS adsorption circuits.

7.2 Batch Adsorption Experiments

Batch adsorption tests for the extraction of platinum, palladium and gold by activated carbon from dilute cyanide solutions were thus conducted in order to determine the kinetic constants, by using the methods explained in the previous section, to enable the modelling of the adsorption circuits in each flowsheet. Due to cost and time constraints similar tests for ion exchange resins were not performed, however, such data was available from literature. The experimental procedure followed and results obtained are discussed in the next sections.

7.2.1 Materials and Methods

Cyanide heap leach solutions were not readily available for use in the adsorption tests, thus synthetic solutions were prepared from precious metal salts that resembled typical cyanide leach liquors as reported by (Mwase, 2009).

Synthetic solutions containing $Pt(II), Pd(II)$ and $Au(I)$ were prepared by dissolving the required amounts of the corresponding metal cyanide salts, $K_2Pt(CN)_4, K_2Pd(CN)_4$ and $KAu(CN)_2$, in alkaline $NaCN$ $(KSCN)$ buffered solutions. The solutions were diluted to obtain the necessary concentrations. The pH of the solutions were adjusted by the use of either $NaOH$ $(1M)$ or H_2SO_4 $(1M)$. All chemicals used were of analytical grade.

Two solutions were prepared as duplicate tests were performed. Table 7.2 shows the target and actual metal concentrations in the solutions prepared, as analysed by ICP-MS.

Table 7.2: Composition of the synthetic cyanide solutions prepared

Element	Concentration (mg/L)	
	Actual (Test 1)	Actual (Test 2)
Pd	12.41	12.43
Pt	0.42	0.43
Au	1.83	1.84

The batch adsorption tests were performed using the traditional bottle-on-rolls method. These specific tests were selected as it has been stated (Fleming *et al.*, 2011) that the kinetic performances achieved with these tests are quite similar to that achieved in large industrial CIP tanks.

Bottles with a total volume of 2.5 litres containing 500 ml of solution in contact with 10 g of activated carbon (at a concentration of 20 g/L) were used. This specific carbon concentration was selected as the AAC Pump Cell mode of operation in CIP plants utilises carbon concentrations between 15 and 60 g/L depending on the ore to be treated (Rogans & McArthur, 2002; Dippenaar & Proudfoot, 2005). The pH of the solution was adjusted to 9.5 prior to the start of each test. This has been shown to be the optimum pH for PGM adsorption (Mpinga, 2012). Tests were performed at 25°C.

Tests were performed for a total of 24 hours at a rotational speed of 105 rpm to ensure that pseudo-equilibrium was achieved. From similar tests conducted by Mpinga (2012) this time was found to be sufficient to achieve this goal. It has been shown (Nicol *et al.*, 1984*a*; van Deventer *et al.*, 1984) that true equilibrium between gold cyanide in solution and activated carbon is still not achieved after several weeks of adsorption. The rolling speed was selected at the maximum of the equipment used, as it has been reported (Fleming *et al.*, 2011) that the rate of adsorption in bottle-on-rolls tests is insensitive to roller speed when bottles without baffles and a smooth inside are used.

Sampling of the solutions was performed at pre-determined time intervals where-after it was filtered and the filtrate sent for analysis by ICP-MS.

7.2.2 Batch Adsorption Test Results

The data obtained from the batch adsorption tests conducted on the extraction of PGMs from dilute cyanide solutions using activated carbon is presented in Table 7.3. It can be seen that the majority of all three metals were extracted within 2 hours where-after extraction was slow due to the low driving force. This is consistent with the findings of Mpinga (2012). The initial rate of adsorption was thus fast and can be attributed to the fact that a large amount of carbon was used and the high mixing efficiencies of the rolling bottle tests. The solutions also did not contain any impurities or fouling elements that could have decreased the rate of adsorption.

Table 7.3: *Batch adsorption test results for the extraction of platinum, palladium and gold from a dilute cyanide solution by activated carbon*

Time (h)	*% Recovery (Test 1)*			*% Recovery (Test 2)*		
	Pd	***Pt***	***Au***	***Pd***	***Pt***	***Au***
0.00	0.00	0.00	0.00	0.00	0.00	0.00
0.25	90.35	90.48	95.69	92.08	92.22	98.27
0.50	96.81	97.46	99.17	97.50	98.02	99.67
0.75	98.80	99.12	99.71	99.02	99.26	99.87
1.0	99.37	99.55	99.82	99.47	99.57	99.90
1.5	99.70	99.77	99.89	99.72	99.79	99.94
2	99.70	99.73	99.87	99.80	99.85	99.96
3	99.81	99.83	99.92	99.86	99.89	99.97
6	99.88	99.86	99.94	99.90	99.91	99.98
12	99.93	99.93	99.96	99.93	99.93	99.97
24	99.94	99.91	99.95	99.94	99.94	99.97

Table 7.4 presents data obtained during batch adsorption tests for the extraction of platinum, palladium and gold from dilute cyanide solution using the ion exchange resins Amberlite PWA 5 and Minix as reported by Schoeman (2012).

Table 7.4: *Batch adsorption test results for the extraction of platinum, palladium and gold from a dilute cyanide solution by Amberlite PWA 5 and Minix (After Schoeman, 2012)*

Time (h)	*% Metal Recovery*			
	Amberlite PWA 5			*Minix*
	Pd	***Pt***	***Au***	***Au***
0	0.00	0.00	0.00	0.00
1	52.31	52.40	33.91	45.55
2	74.42	75.13	52.27	62.35
3	84.82	85.12	60.77	73.39
4	90.10	90.41	64.71	78.67
8	98.00	98.44	80.24	90.51
16	99.34	99.42	87.46	95.65
24	99.50	99.64	92.30	96.83
36	99.50	99.82	93.82	97.32
48	99.50	99.95	94.90	97.97

It can be seen that Amberlite PWA 5 readily extracted all metals from solution. The reason that gold is the only metal reported here for the extraction by Minix is that Minix is a gold selective resin that rejects platinum and palladium during continuous contacting (Schoeman, 2012; Schoeman *et al*., 2012). The kinetics thereof was thus not determined.

The data presented in Table 7.4 were the only data available for the extraction of PGMs with the use of ion exchange resins from dilute cyanide solutions and were thus used in the subsequent determination of the respective kinetic constants of platinum, palladium and gold.

7.3 Determination of Kinetic Constants

Table 7.5 shows the kinetic constants obtained from a non-linear least squares fit of equation (8) to the batch adsorption test data presented in Table 7.3 for the activated carbon tests. The averages of the two tests are presented here. The batch kinetic model was found to fit the data well as indicated by the coefficient of determination (R^2) of close to 1 for all metals. As no kinetic data for the adsorption of platinum and palladium from dilute cyanide solutions onto activated carbon have been reported in literature, a comparison with data reported for gold was used to provide a sanity check. Nicol *et al.* (1984*a*) reported kinetic constants (kK) of 548 h^{-1} and 555 h^{-1} for gold obtained during similar batch adsorption tests performed on Grootvlei ore. The higher value obtained in this study could be explained by the fact that smaller bottles were used in the batch adsorption tests with higher mixing efficiencies resulting in an increase in the kinetic rate constant improving the overall rate of adsorption (Nicol *et al.*, 1984*b*; 1984*c*). Furthermore a synthetic solution was used that did not contain any base metals or other impurities, such as silica, that have a negative effect on the rate of adsorption. Taking this into account, the values for the kinetic constants were deemed acceptable.

Table 7.5: *Kinetic constants for platinum, palladium and gold for activated carbon*

Activated Carbon				
Metal	k *(h^{-1})*	K	kK *(h^{-1})*	R^2
Pd	0.047	10 350	490	0.996
Pt	0.037	13 511	494	0.998
Au	0.021	30 661	657	0.999

Table 7.6 shows the kinetic constants obtained from a non-linear least squares fit of equation (8) to the batch adsorption test data presented in Table 7.4 for ion exchange resins.

The model, developed for the adsorption by activated carbon, was found to fit these data very well as indicated by the coefficient of determination (R^2) values of close to 1 for all metals. Kinetic constants for platinum and palladium extracted by Minix were not determined due to the fact that the high affinity of the Minix resin for gold resulted in the platinum and palladium being kicked off as soon as it were loaded onto the resin (Schoeman, 2012).

Table 7.6: *Kinetic constants for platinum, palladium and gold for Amberlite PWA 5 and Minix ion exchange resins*

	Amberlite PWA 5				*Minix*			
Metal	k (h^{-1})	K	kK (h^{-1})	R^2	k (h^{-1})	K	kK (h^{-1})	R^2
Pd	0.013	84 427	1 072	0.993	-	-	-	
Pt	0.011	96 362	1 083	0.993	-	-	-	
Au	0.051	10 642	548	0.956	0.024	29 594	725	0.967

In order to access the likely applicability of the model and associated kinetic constants it was necessary to review the mechanism of the rate of adsorption of gold onto ion exchange resins and activated carbon. Nicol *et al.* (1984*a*) have stated that the rate of extraction of gold cyanide by ion exchange resins and activated carbon is controlled by film diffusion in the initial stages (less than 30% of the gold equilibrium loading) and by both film and intra-particle diffusion in the latter stages, as equilibrium is approached. The respective rates of extraction by resins and carbon will thus be similar during the initial stages, and will only differ as a result of the difference in the size of the resin beads (typically 0.85 - 1 mm in diameter) and the carbon granules (1 - 3 mm). As typical gold CIP plants operate well below equilibrium (Nicol *et al.*, 1984*a*) the rate of extraction by ion exchange resins is expected to be about twice the rate of extraction by activated carbon as, under film diffusion-controlled conditions, the rate of loading is inversely proportional to the mean particle-diameter of the adsorbent (Kotze *et al.,* 2005).

By this definition the kinetic constants obtained for platinum and palladium by extraction with Amberlite PWA 5 can be interpreted as meaningful as they are approximately double those obtained for the extraction by activated carbon. The reason for the lower kinetic constant with respect to the extraction of gold using Amberlite PWA 5 is due to the fact that this resin has a higher affinity for the divalent cyanide ions ($Pt(CN)_4^{2-}$, $Pd(CN)_4^{2-}$) as opposed to the monovalent aurocyanide ion ($Au(CN)_2^-$) due to its larger functional group (Schoeman, 2012). The gold thus had to compete directly with the platinum and palladium for active sites.

For the Minix resin the gold kinetic constant was, although higher than that for carbon, still relatively lower than what would be expected for a high gold selective resin. This may have been due to the low concentration of gold in solution used in the batch adsorption tests conducted by Schoeman (2012) and also due to the very low resin to solution ratio (1:1000) used in order to enhance the competitive loading characteristics.

It is a well-known fact that feed solution concentration has an effect on the kinetics of adsorption (Nicol *et al*., 1984*c*). The values calculated here were, however, used in the subsequent modelling of the adsorption circuits as they were based on the only data available for the extraction of gold via the Amberlite PWA 5 and Minx ion exchange resins from dilute cyanide solutions containing extensive amounts of base metals as well as platinum group metals in concentrations that could be expected of a typical cyanide heap leach on Platreef ore.

7.4 Adsorption Optimization

The use of the Nicol-Fleming model in the optimization of an adsorption circuit is illustrated in the next section. Some design considerations are presented, followed by a discussion on the different operating variables that are used for optimization. Lastly the optimization procedure used to arrive at an optimum design is presented. This is a review of the procedure outlined by Nicol *et al.* (1984*b*).

7.4.1 Design Considerations

In the design procedure several considerations were made to ensure practically unrealistic designs were excluded from the optimisation process. This included limits on the number of adsorption stages, upgrade ratio and carbon concentration is each tank.

These considerations are discussed in further detail below:

- The number of adsorption stages should not be more than 10. The cost benefits of operating at or above this amount become diminished and the relative performance expected of such a circuit could be poor, as carbon fouling and carbon attrition become severe (Stange, 1991, 1999).

- The design loading and subsequently the upgrade ratio should adhere to general plant data such as that presented in Figure 2.4. If the optimum design was found to be below this general trend it was deemed satisfactory, while it was limited if the optimum design was found to be above this trend as there are several practical implications of operating at too high upgrade ratios, as will be discussed in the next section. Typical upgrade ratios for plants utilising the AAC Pump Cell mode of operation is in the range of 1 200 - 2 500 (Whyte *et al.*, 1990; Rogans & McArthur, 2002).
- The adsorbent concentration per stage should not exceed 60 g/L. This is the maximum operating value used in adsorption circuits utilising the AAC Pump Cell mode of operation (Whyte *et al.*, 1990; Rogans & McArthur, 2002). Furthermore, as it has been discussed, the model under-predicts the amount of adsorbent required. Thus if an optimum were found to be close to this 60 g/L limit it could lead to detrimental performance if the plant would be commissioned at these conditions.

7.4.2 Optimization Parameters

The design of an adsorption circuit involves around many important decisions regarding the operation of the plant. Most importantly these decisions include the desired extraction efficiency (i.e. the barren solution concentration), the target gold on carbon loading, and the number of adsorption stages required to achieve this metallurgical performance.

These different variables are all inter-dependent on one another and significant trade-offs between them and the effects that they have are required to arrive at an optimal design. The trade-offs that these variable have on the overall performance and operation of the circuit are discussed in further detail in the next few sections.

7.4.2.1 Effect of changing the desired barren solution concentration

Increasing gold recovery, thus decreasing the barren solution concentration, requires that larger carbon inventories be employed, increasing gold lockup. Furthermore carbon needs to be moved more rapidly throughout the plant to ensure effective carbon activity, reducing the gold on carbon loadings, increasing the size of the elution and regeneration circuits, negatively impacting process economics (Nicol *et al.,* 1984*b*; Stange, 1991, 1999). An increase in overall recovery, however, increases the total income of the process. It can thus be seen that there are significant trade-offs between selecting the desired barren solution concentration.

7.4.2.2 Effect of changing the target gold on carbon loading

Increasing the gold on carbon loading leads to less carbon being transferred to the elution and regeneration circuits, decreasing capital and operating costs (Stange, 1991, 1999). However, this requires longer carbon residence times that lead to carbon fouling and excessive fines generation due to attrition (Bailey, 1987). There is, however, a smaller carbon loss in the rest of the plant due to decreased carbon handling (Staunton, 2005). Furthermore the associated gold locked up with larger inventories and higher loadings negatively impacts the process economics (Nicol *et al.*, 1984*b*).

7.4.2.3 Effect of changing the number of adsorption stages

Increasing the number of adsorption stages results in the carbon inventory in each tank being lowered, reducing gold lockup, if the same metallurgical performance, i.e. barren concentration, is too be achieved (Bailey, 1987; Nicol *et al.*, 1984*b*). This trend, however, flattens off as the number of stages is further increased (at about 8 - 10 stages), as effective contact time between the solution and adsorbent is required. Typical CIP plants operate at around 6 - 8 stages (Stange, 1999; Staunton, 2005). Higher gold on carbon loadings can be achieved if the adsorption cycle is kept constant, which has several cost implications. Higher loadings lead to less carbon being transferred to the elution and regeneration circuits, decreasing the capital and operating costs (Stange, 1991, 1999).

Caution should be taken in following this approach as it is known that an increase in the total number of stages, while keeping the adsorption cycle constant, leads to an increase in the average carbon residence time required, which could lead to excessive carbon fouling and breakage due to attrition (Stange, 1999).

7.4.3 Optimization Procedure

From the above discussion it can be seen that there are significant trade-offs between the different operating variables. The next section shows how these variables together with the derived model can be used to arrive at an optimum design, keeping in mind the aforementioned design considerations.

The model, with the use of the kinetic constants, as determined in section 7.3, can be combined with an economic model to allow for the optimization of the adsorption circuits.

In this case the model was directly applied to the flowsheet mass balance that was linked to a costing algorithm. This enabled the net present value (NPV) of the project to be calculated as the different design decisions were made (please refer to Chapters 5 and 6 for flowsheet considerations and process economic discussions).

The optimization procedure that was used is an iterative method of obtaining the optimum design. This is due to the interdependence of the different variables. This method of optimization is given below (Nicol *et al.*, 1984*b*):

- An assumption as to the number of stages necessary to obtain a certain barren solution concentration is made. The optimum loading is then calculated with the use of the model combined with the mass balance and the economic algorithm.
- This optimum loading is then used in the calculation of the optimum number of stages followed by the estimation of the optimum extraction efficiency.
- If the assumed values were to be correct then an optimized design was found. If the optimum values obtained did not agree with those assumed at the start, a new cycle of calculations were performed using these optimum values as the new starting values.
- This procedure was repeated until the respective values converged.

To illustrate this optimization approach, results from the CIS 1 flowsheet are presented and serves as an example as to how this model can be used in a cost-minimization procedure. It should be noted that optimization was based on the metal with the slowest kinetics.

For the combined and primary leach solutions this was found to be palladium, as its respective concentration in solution and subsequent stream value was the highest and kinetics the slowest. It was thus assumed that the amount of carbon and time required to extract the palladium would be sufficient for the extraction of both gold and platinum.

As platinum was the only metal present in the secondary leach solution, it was subsequently used in the optimization of that stream of the adsorption circuit.

For this example it was assumed that 7 adsorption stages were required to produce a barren solution concentration of 0.037 mg/L for palladium. Various loadings were then entered into the model from which an optimum loading was obtained. From Figure 7.4 it can be seen that the optimum loading was found to be 3 774 g/t. This optimum exists due to the various factors discussed in section 7.4.2.

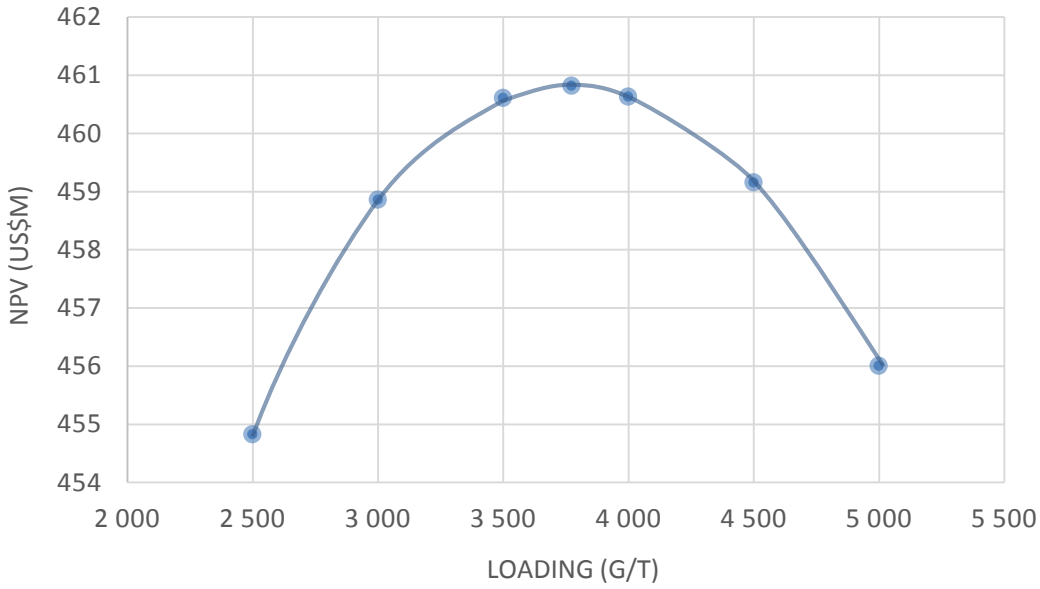

Figure 7.4: *Optimization of the Carbon Loading*

This optimum loading, together with the barren solution concentration assumed, was then used to estimate the optimum number of adsorption stages, as presented in Figure 7.5. It can be seen that the assumed value of 7 stages was correct.

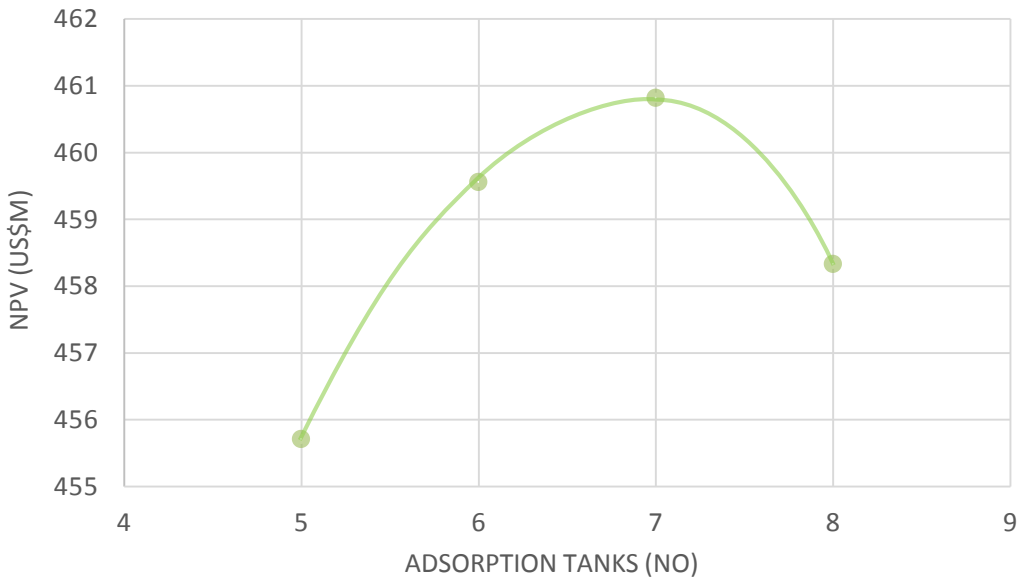

Figure 7.5: *Optimization of the Number of Adsorption Stages*

Finally the estimated optimum loading and number of adsorption stages were used in estimating the optimum barren solution concentration. From Figure 7.6 it can be seen that the assumed value of 0.037 mg/L was correct. If this value were found to be different from that assumed, then the procedure would have been repeated using the new values until they converged.

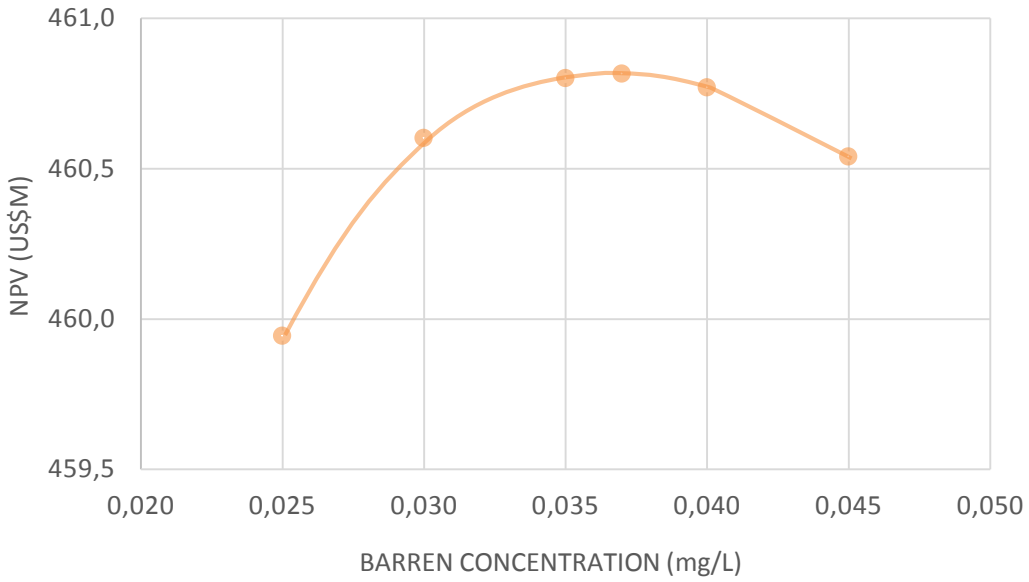

Figure 7.6: *Optimization of the Extraction Efficiency*

The whole procedure was presented thus showing how this model combined with economic models can be used in finding an optimum design for the adsorption circuit and what a change in the different parameters could do to the potential economic outcome of the project.

Chapter 8

Simulation Results and Flowsheet Comparison

The simulations of each of the different flowsheet alternatives were performed with the use of the data and assumptions as discussed in Chapter 5. This allowed for cost estimations to be performed that, when combined with the mass balances, were used in the subsequent flowsheet optimization procedures, as discussed in Chapters 6 and 7 respectively. All simulation and optimization was performed with the use of Microsoft Excel.

In this chapter the results obtained from these simulations are discussed. Mass balance and operating data for each of the flowsheets options are discussed. Detailed comparisons between the different flowsheet alternatives, based on the produced products and several financial parameters, are then presented. The optimum process alternative are identified and discussed.

8.1 Simulation Results

8.1.1 SART

The SART plants for all flowsheet alternatives were similar, being either a SART plant treating the combined solution (CIS 1, RIS 1 & 3) or two SART plants treating the leach solutions separately (CIS 2, RIS 2, 4 & 5). Results of these three plants are discussed below and include the composition of the feed and discharge streams and the base metal precipitates produced.

The SART plant feed and discharge stream compositions are given in Table 8.1. With the assumption of 99% base metal (Cu, Ni and Zn) removal, the base metal concentrations of the discharge streams (feed to the adsorption circuits) have been reduced to less than that of the precious metal content, except in the case of the very dilute secondary leach solution. It is believed that these reductions in base metal concentration will have significant benefits in the performances of commercial adsorption and elution circuits. The slight decrease in the concentration of the precious metals was as a result of the chemical additions, mainly pH adjustments, and the filter wash water recycled back to the processes.

Table 8.1: *SART feed and discharge stream metal concentrations*

	Combined SART		Primary SART		Secondary SART	
	Feed	Discharge	Feed	Discharge	Feed	Discharge
Flow (t/h)	353.37	358.54	58.97	60.15	294.39	298.39
	Concentration (mg/L)					
Cu	102.30	1.00	351.33	3.35	52.42	0.51
Ni	99.32	0.97	266.38	2.54	65.85	0.65
Zn	3.10	0.03	7.97	0.08	2.13	0.02
Fe	0.26	0.20	0.52	0.40	0.21	0.16
Pt	1.17	1.15	2.55	2.48	0.89	0.88
Pd	2.23	2.20	13.38	13.06	0.00	0.00
Au	0.32	0.31	1.90	1.86	0.00	0.00

Table 8.2 shows the flowrates and compositions of the base metal precipitate products. The precipitates consisted mainly of copper sulphide and nickel sulphide, due to the excessive amounts of these metals present in the leach solutions. The PGM compositions of these precipitates, as a result of the moisture content, were less than 0.01% or about 1 mg/L.

Table 8.2: *Base metal sulphide precipitate products*

	Combined SART	Primary SART	Secondary SART
Flow (kg/h)	108.53	55.20	53.33
	Composition %		
Cu_2S	48.89	55.10	42.47
NiS	49.51	43.57	55.65
ZnS	1.49	1.26	1.73
FeS	0.11	0.07	0.15
$PGM's$	<0.01	<0.01	<0.01

The revenue from selling these precipitates is low compared to the PGM and gold products produced, as will be discussed later. The revenue from the base metal and gypsum precipitates could, however, allow for the capital costs of the SART plants to be paid back within 4 years.

The overall performance of the different SART plants was found to be significant with regards to the predicted PGM precipitates produced. A 1% decrease in the overall base metal recoveries resulted in an increase of approximately 15% in the base metal content of the PGM precipitates, due to the very high base metal content in the leach liquors. It is thus of utmost importance that the SART plants are operated at the highest base metal recoveries possible.

8.1.2 Carbon Adsorption

The optimum designs of the carbon adsorption sections for each of the CIS flowsheet alternatives, as obtained with the use of the adsorption model, are presented in Table 8.3.

Table 8.3: *Carbon adsorption section designs*

	CIS 1	*CIS 2*	
	Only Circuit	*Primary Circuit*	*Secondary Circuit*
PLS Flow (kg/h)	358.54	60.15	298.39
Tanks (operating)	7 (8)	7 (8)	7 (8)
Size (m^3)	122	21	102
Carbon Conc (g/L)	12.41	13.48	11.6
Carbon Inventory (t)	10.40	1.89	8.09
Carbon Consumption (g/t)	40	70	32

The CIS 1 option required only one adsorption section, as the leach solutions were combined to produce a single feed. The optimum number of operating stages required was estimated to be 7, with an extra tank required as a result of the carousel mode of operation. Tank sizes were estimated to be 122 m^3, based on the assumed 20 minute solution residence time.

The carbon concentration in each tank, calculated with the use of the model, was low, if compared with typical pump cell operations which operate at between 20 - 60 g/L (Whyte *et al*., 1990; Stange 1999; Rogans & McArthur 2005; Dippenaar & Proudfoot, 2005). This has been shown to be a problem with the model as it under predicts the required amount of carbon, when compared to actual plant performance data (Nicol *et al*., 1984*b*). As a result the associated value material lock-up was also low. These values were accepted, however, as the contributions of these parameters to the overall process economics were minimal, compared to the overall capital and operating costs.

The CIS 2 option required two adsorption sections as the cyanide leach solutions were treated separately. Smaller tanks were thus required, with the primary circuit operating with 7 stages, while 7 stages were also found to be the optimum for the secondary circuit. Carbon concentrations and the value material lock-up were also estimated to be low in these two circuits, with the reason already mentioned above.

The required number of adsorption stages compared well to typical gold carbon adsorption plants treating similar tonnages, which utilize between 6 - 8 stages (Stange, 1999; Rogans & McArthur, 2002). Carbon consumption rates also compared well to values reported in literature. The lower solution throughput and higher grades of the CIS 2 primary circuit resulted in lower upgrade ratios, as discussed later. Faster carbon movement was thus required which led to higher carbon consumption rates.

Operating data for the different adsorption circuits are presented in Table 8.4. All three adsorption circuits showed high overall recoveries due to the fast adsorption kinetics and the fact that the elution of all three metals were effective, resulting in relatively low recycled carbon loadings. The low recycled loadings allow for the effective extraction of the value minerals, even at the low concentrations of the feed solutions to the final adsorption tanks. The driving forces behind the adsorption mechanisms were still sufficient to reduce the barren concentrations too low values.

Table 8.4: *Carbon adsorption circuit operating data*

	CIS 1	*CIS 2*	
	Only Circuit	*Primary Circuit*	*Secondary Circuit*
PLS Flow (t/h)	358.54	60.15	298.39
Carbon Flow (t/h)	0.209	0.065	0.138
Feed Conc (mg/L)			
Pt	1.15	2.48	0.88
Pd	2.20	13.06	-
Au	0.31	1.86	-
Carbon Loadings (g/t)			
Pt	1 973	2 317	1 890
Pd	3 774	12 211	-
Au	544	1 747	-
Recycled Loadings (g/t)			
Pt	28	32	26
Pd	70	228	-
Au	10	33	-
Recovery (%)			
Pt	98.88	99.46	98.35
Pd	98.38	99.27	-
Au	99.69	99.84	-
Upgrade Ratio	1 719	935	2 125

Estimated recoveries of the precious metals in the primary section of CIS 2 were higher than that observed for CIS 1. This was mainly due to the higher feed concentrations, increasing the driving force behind the adsorption mechanisms. Although the barren concentrations of the primary solution leaving the CIS 2 adsorption circuit were higher, the lower solution flow rate, higher feed concentration, and faster kinetics resulted in higher overall recoveries. A slightly lower platinum recovery was estimated for the secondary CIS 2 adsorption circuit, compared to that achieved in CIS 1, as the feed solution concentration was very low, increasing the difficulty of recovering the platinum.

In comparing the upgrade ratios of the three circuits, the estimated values compared very well to typical gold adsorption circuit operating data at similar feed concentrations, as discussed in section 2.5.4 with particular reference to Figure 2.4. Upgrade ratios typically decrease with increasing feed solution concentration. The higher concentrations result in the carbon being loaded to higher values, which greatly reduces the driving force behind the adsorption mechanisms. Thus lower upgrade ratios are used to maintain fast adsorption kinetics whilst achieving good recoveries.

The estimated overall operation and performance of the carbon adsorption circuits were thus deemed to provide an accurate representation of what could be expected if actual plants would be built, based on the above discussion and comparison to literature values.

8.1.3 Carbon Elution and Final Metal Recovery

Carbon loadings prior and after elution for the two carbon flowsheet alternatives are presented in Table 8.5. The AARL elution method was used in all circuits. The assumption was made that only one elution is performed per day, and the elution circuits were thus sized according to the amount of carbon transferred to the elution column per day.

The CIS 1 elution circuit was thus sized to be able to treat 5 tons of carbon, while the primary and secondary columns of CIS 2 were smaller at about 1.6 and 3.3 tons respectively. The combined size of the two separate circuits was the same as that of the CIS 1 circuit, as would be expected.

It can be seen from Table 8.5 that the total base metal loadings were less than those of the precious metals, mainly due to the benefits of installing the SART plants. All metal loadings were estimated to be reduced to low values by the elution with de-ionized water.

Acid washing removed about 50% of the base metals loaded onto the carbon, while the subsequent elution step stripped the carbon of almost all base metals. Low precious metal loadings on the eluted carbon, as a result of the high elution efficiencies that can be achieved with de-ionized water, allowed for the effective extraction of all these metals in the adsorption sections. The higher eluted carbon loadings of the CIS 2 primary circuit were due to the high loadings achieved during adsorption and the fact that fixed elution percentages were assumed.

Table 8.5: *Carbon loadings before and after elution*

	CIS 1		*CIS 2*			
Plant Section	*Only Circuit*		*Primary Circuit*		*Secondary Circuit*	
Elution Method	AARL		AARL		AARL	
Carbon Eluted (t/d)	5.0		1.6		3.3	
Loadings (g/t)	Loaded	Eluted	Loaded	Eluted	Loaded	Eluted
Cu	462	37	837	67	300	24
Ni	1 523	30	2 154	43	1 278	25
Zn	42	1	57	1	36	1
Fe	119	91	128	97	121	92
Pt	1 973	30	2 317	35	1 890	28
Pd	3 774	75	12 211	244	-	-
Au	544	11	1 747	35	-	-

Overall adsorption-elution circuit performances were thus good, as elution of the loaded carbon was sufficiently effective that high driving forces were obtained in the adsorption section, leading to the effective extraction of the precious metals during adsorption.

The resulting eluate flowrates and concentrations are presented in Table 8.6 and Table 8.7. The pregnant eluate concentrations were high, which resulted in more concentrated solutions compared with the pregnant leach liquors. These upgraded solutions compared well to the 2 - 3 fold upgrade in gold concentrations achieved in industry (Fleming, 1992). Higher loaded carbon values resulted in more concentrated solutions, even though longer elution times were required to achieve the same overall elution efficiencies. In upgrading the solution to the more concentrated ones, recovery of the metals directly from these solutions becomes possible, both from a metallurgical and economic standpoint.

These upgraded solutions were sent to thermal hydrolysis where the bulk of the precious metals were precipitated from solution. It can be seen that the concentrations of the base and platinum group metals were reduced to less than 2 mg/L in all three scenarios.

The gold concentrations were still high as a result of the fact that the operating conditions of the autoclaves were selected to be below those at which the aurocyanide complex is destroyed (resulting in precipitation of the gold in its metallic form). These gold values were recovered by electrowinning of the thermal hydrolysis discharge solutions, resulting in low barren gold values (<0.35 mg/L) at an assumed recovery of 99%. As it was assumed that 98% of the remaining base and precious metals would co-deposit, their respective barren concentrations were extremely low. Research is still required to determine if it would be possible for both platinum and palladium to be co-extracted with gold, as it would increase the value of the gold product significantly.

Table 8.6: *Eluate Concentrations (CIS 1)*

	CIS 1		
Plant Section	*Only Circuit*		
Eluate Required (t/d)	97.19		
Concentration (mg/L)	Pregnant Eluate	After Thermal Hydrolysis	After Electrowinning
Cu	21.76	0.11	0.00
Ni	37.06	0.19	0.00
Zn	1.04	0.01	0.00
Fe	1.16	0.01	0.00
Pt	99.45	0.50	0.01
Pd	189.24	0.95	0.02
Au	27.28	21.83	0.22

Table 8.7: *Eluate Concentrations (CIS 2)*

	CIS 2				
Plant Section	*Primary Circuit*			*Secondary Circuit*	
Eluate Required (t/d)	59.91			50.14	
Eluate Concentration (mg/L)	Pregnant Eluate	After Thermal Hydrolysis	After Electrowinning	Pregnant Eluate	After Thermal Hydrolysis
Cu	19.95	0.10	0.00	18.05	0.09
Ni	26.63	0.13	0.00	39.71	0.20
Zn	0.72	0.00	0.00	1.15	0.01
Fe	0.63	0.00	0.00	1.50	0.01
Pt	59.09	0.30	0.01	121.85	0.61
Pd	309.92	1.55	0.03	0.00	0.00
Au	44.35	35.49	0.35	0.00	0.00

8.1.4 Resin Adsorption

The optimum designs and performance data of the resin adsorption circuits for each of the RIS flowsheet alternatives are presented in Table 8.8.

For all flowsheet alternatives 6 - 7 adsorption tanks were required. These tanks were smaller in size to that of the carbon adsorption circuits, as it was assumed that the improved kinetics of the resins would result in smaller tanks being required. This compared well to the Golden Jubilee gold mine where similar number of resin adsorption tanks were used when its CIP operation was replaced (Fleming, 1992; Kotze *et al*., 2005).

Resin concentrations per tank were low, but as mentioned before (refer to section 8.1.2), the model under predicts the amount required. Due to the improved kinetics and the substantial increase in cost of the resin in comparison with activated carbon, the optimum concentrations and total resin inventories were lower than that of carbon. This is in agreement with findings mentioned by Fleming (1992). Resin losses were estimated to be between 2 - 8 g/t, which compared well to values reported in literature for RIP operations (8 - 10 g/t) keeping in mind that RIS losses is typically 30 - 40% less than in RIP (Fleming, 1992; Marsden & House, 2006).

The estimated loadings of the precious metals were found to be higher than those of the carbon circuits and although resin elution efficiencies were higher, these higher loadings resulted in higher loadings for the resin recycled back to the adsorption circuits.

The overall recoveries of platinum and palladium were found to be very high, at over 97% and 99% respectively. This was mainly due to the fast adsorption kinetics of the Amberlite PWA 5 resin, and the high solution feed grades of palladium and platinum. Furthermore the recycled loadings were relatively low ensuring effective adsorption. The same general trend in the difference of the estimated recoveries that were observed between the single and multiple carbon plant options were seen in the RIS flowsheet options.

The recoveries of gold were, however, significantly lower than those estimated for platinum and palladium, especially in the flowsheets using Amberlite PWA 5 as the adsorbent. The factors believed to have played a major role in these low recoveries were the higher selectivity of the resin for platinum and palladium (Schoeman, 2012), the slower kinetics of gold adsorption compared to that of platinum and palladium and the fact that gold elution, in the case of elution with zinc cyanide, were poor. Elution performance is discussed in the next section.

Table 8.8: Resin adsorption circuit designs

	RIS 1	*RIS 2*		*RIS 3*	*RIS 4*		*RIS 5*	
Plant Section	*Primary*	*Primary*	*Secondary*	*Primary*	*Primary*	*Secondary*	*Primary*	*Secondary*
PLS Flow (kg/h)	358.54	60.15	298.39	358.54	60.15	298.39	60.15	358.54
Feed Conc (mg/L)								
Pt	1.15	2.48	0.88	1.15	2.48	0.88	2.48	1.15
Pd	2.20	13.06	-	2.20	13.06	-	13.06	2.19
Au	0.31	1.86	-	0.31	1.86	-	1.86	0.02
Tanks (operating)	7 (8)	6 (7)	6 (7)	7 (8)	6 (7)	6 (7)	7 (8)	6 (7)
Size (m^3)	60	10	50	60	10	50	10	60
Resin Employed	Amberlite	Amberlite	Amberlite	Amberlite	Amberlite	Amberlite	Minix	Amberlite
Resin Conc (g/L)	8.09	10.34	5.60	7.61	13.44	5.60	5.10	8.09
Resin Losses (g/t)	2.71	6.30	2.11	3.55	7.58	2.11	4.45	2.03
Metal Loadings (g/t)								
Pt	4 850	4 004	4 485	6 671	7 032	4 485	150	6 376
Pd	9 100	20 631	-	12 407	35 907	-	62	11 921
Au	1 527	4 004	-	1 564	4 963	-	10 402	24
Recycled Loadings (g/t)								
Pt	144	119	133	196	207	133	147	189
Pd	90	204	-	1	4	-	1	118
Au	529	1 387	-	5	17	-	102	8
Recovery (%)								
Pt	99.60	99.74	97.90	99.44	99.82	97.90	0.02	99.10
Pd	99.66	99.76	-	99.58	99.90	-	0.08	99.14
Au	77.64	89.83	-	87.91	96.72	-	98.6	64.88
Upgrade Ratio	4 229	1 646	5 105	5 641	2 753	0 5 105	5 599	5 483

For the first two flowsheet options, RIS 1 & RIS 2, elution was performed with the use of zinc cyanide. Based on the elution recoveries reported by Schoeman (2012), gold elution was poor at only about 65% recovered within 15 bed volumes. This led to large amounts of gold still present on the resins recycled back to the adsorption circuits. These high recycled loadings, and the fact that the gold grade of the solutions was low, resulted in low driving forces for adsorption that in turn resulted in low recoveries. The gold recovery in RIS 2 (89.83%) was, however, significantly higher than that achieved in RIS 1 (77.64%), mainly due to the considerably higher gold feed grade of the primary circuit.

The gold recoveries were estimated to be much higher in the RIS 3 and RIS 4 flowsheet options. Although Amberlite PWA 5 was still the adsorbent, the fact that a sequential elution technique was performed, in which gold and palladium were stripped with acidic thiourea followed by platinum elution with zinc cyanide, resulted in very low recycled gold loadings. The adsorption reactions thus had a higher driving force which resulted in higher extraction efficiencies.

The overall gold recovery of the RIS 5 option was also high, due to the gold selective Minix resin (with improved gold adsorption kinetics) used followed by the Amberlite PWA 5 resin for extraction of the platinum, palladium and gold not extracted in the primary circuit. Gold extracted with Amberlite was low due to the very low feed grade of the combined solution.

The estimated upgrade ratios were found to be about 2 - 4 times those estimated in the carbon adsorption circuits. Upgrade ratios varied from 1 646 for the higher grade solutions to 5 641 for the lower grade solutions. Data from literature were used to assess the credibility of these estimated values. Data from the Golden Jubilee gold mine, which replaced its CIP plant with a RIP plant, were used for this purpose. In treating a gold solution tenor of between 1 - 1.5 g/t, the observed carbon upgrade ratios were about 800 - 1 500. These upgrade ratios, however, increased to about 4 000 - 6 000 when the RIP plant was in full operation (Fleming, 1993; Kotze *et al*., 2005). The RIP upgrade ratios were thus 4 times those achieved during CIP operations. The upgrade ratios obtained from the simulations in this study thus compared well with those achieved at the Golden Jubilee gold mine.

Taking all of the above into consideration and the fact that similar data were not available for platinum and palladium, the simulation results were deemed to provide a reasonable representation of what could be expected in full-scale resin-in-solution operations.

8.1.5 Resin Elution and Final Metal Recovery

Resin loadings prior and after elution for the five resin flowsheet alternatives are presented in Table 8.9. Two different elution methods were used in the different flowsheets, and included the elution with zinc cyanide, acidic thiourea or a combination thereof. The same assumption was made that only one elution is performed per day. The elution circuits were thus sized according to the amount of resin transferred to the elution column per day. As a result of the higher upgrade ratios, the elution sections were significantly smaller to that of the CIS circuits.

It can be seen in Table 8.9 that the loadings of all the metals were far larger to those achieved during carbon adsorption. Even though the base metal recoveries were lower, the fact that less resin was moved per day resulted in significantly higher base metal loadings.

The bulk of the nickel and zinc loaded onto the resins (98%) was, however, stripped prior to elution, with the use of an acid washing step. This resulted in much lower base metal loadings during precious metal elution that increased the grade of the PGM precipitates, as discussed in the next section. Although the bulk of the zinc was removed during acid washing of the resins, the use of zinc cyanide as the eluting agent resulted in a high loading of 1 000 g/t, even after regeneration, recycled back to the adsorption circuits.

It can be seen that the flowsheets that used zinc cyanide as the only eluting agent (RIS 1 & RIS 3) had high residual gold loadings recycled back to the adsorption sections. As explained above, this had a detrimental effect on overall gold adsorption efficiencies that will have negative effects on the overall process economics.

In the flowsheets that used the sequential elution method (RIS 3 & RIS 4), it can be seen that all metals were almost completely stripped from the resins. Platinum loadings were, however, high due to the ineffective elution thereof by acidic thiourea. The gold loadings on the eluted resins were very low, and resulted in the high adsorption efficiencies of gold in the subsequent adsorption circuits. This could prove to hold great benefits with regards to the process economics. The costs of the extra elution cycle would on the other hand reduce these benefits that could be expected.

These economic scenarios are discussed in detail in the next sections.

Table 8.9: Resin loadings before and after elution

Plant Section	*RIS 1*		*RIS 2*				*RIS 3*		
	Primary		*Primary*		*Secondary*		*Primary*		
Elution Method	Zinc Cyanide		Zinc Cyanide		Zinc Cyanide		Acid Thiourea		Zinc Cyanide
Resin Eluted (t/d)	2.09		0.92		1.41		1.52		
Loadings (g/t)	Loaded	Eluted	Loaded	Eluted	Loaded	Eluted	Loaded	Eluted	Eluted
Cu	830	8	1 062	11	526	5	1 135	227	2
Ni	3 204	12	3 101	13	2 626	10	4 405	29	6
Zn	1 111	1 000	1 106	1 000	1 093	1 000	1 147	16	1 000
Fe	42	1	32	1	42	1	58	54	1
Pt	4 850	146	4 004	120	4 485	135	6 671	6 671	200
Pd	9 100	91	20 631	206	-	-	12 407	124	1
Au	1 527	535	4 004	1 401	-	-	1 564	16	5

Plant Section	*RIS 4*					*RIS 5*			
	Primary			*Secondary*		*Primary*		*Secondary*	
Elution Method	Acid Thiourea		Zinc Cyanide	Zinc Cyanide		Acid Thiourea		Zinc Cyanide	
Resin Eluted (t/d)	0.52			1.41		0.26		1.58	
Loadings (g/t)	Loaded	Eluted	Eluted	Loaded	Eluted	Loaded	Eluted	Loaded	Eluted
Cu	1 849	370	4	526	5	1 172	234	1 057	11
Ni	5 603	38	8	2 626	10	1 438	9	4 025	15
Zn	1 184	16	1000	1 093	1 000	277	4	1 106	1 000
Fe	56	52	1	42	1	758	705	55	1
Pt	7 032	7 032	211	4 485	135	150	150	6 376	191
Pd	35 907	356	4	-	-	62	1	11 921	119
Au	4 963	50	17	-	-	10 102	104	24	8

As a result of the much higher loadings observed for the RIS flowsheets compared to the CIS flowsheets, the subsequent eluate concentrations were found to be 2 - 3 times higher. This could prove to have significant benefits in final metal recovery, as it may be easier to recover the metals from higher grade solutions. The impact thereof is, however, unknown at present but should certainly be investigated as it could be a major consideration in selecting between a RIS or CIS circuit.

8.1.6 Overall Recovery

The overall precious metal recoveries of the different flowsheet alternatives are presented in Table 8.10. It can be seen that, in the case of platinum and palladium, the RIS flowsheets achieved higher overall recoveries. This was mainly due to the enhanced adsorption kinetics, higher elution efficiencies and the fact that less value material was lost due to decreased fines generation, when compared to CIS.

The gold recoveries were, however, significantly higher for the two CIS flowsheet options. As explained earlier, the higher selectivity of the resins for platinum and palladium and the ineffective elution of gold form the resins, resulted in the low recoveries observed for the RIS flowsheets. The gold selective Minix resin (RIS 5), however, proved to be effective in recovering high amounts of gold.

Table 8.10: *Overall precious metal recovery of the different process alternatives*

	CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
	Overall Recovery (%)						
Pt	97.68	97.21	99.31	97.94	98.61	97.76	98.23
Pd	97.16	98.11	99.39	99.22	99.29	99.32	98.06
Au	97.67	97.88	76.38	88.16	86.50	94.90	97.34
Total	**97.37**	**97.81**	**97.40**	**97.87**	**97.99**	**98.45**	**98.05**

The gold content in the feed streams to these processes were, however, significantly lower than that of platinum and palladium and only accounted for 8.5% of the total precious metals. Due to this fact the overall precious metal recoveries of the RIS flowsheets were higher than that of the CIS options, even though the gold recoveries were lower, as a result of the higher platinum and palladium recoveries. Based on this it can be concluded that RIS technology could have greater potential in the application thereof to PGM ore deposits with low gold contents.

8.2 Carbon-in-Solution Flowsheet Comparison

The following section is a comparison between the two CIS flowsheet alternatives (CIS 1 and CIS 2). Comparisons are made based on product quality and process economics. The raw data presented in the figures below can be found in Appendix B, C and D.

8.2.1 Product Quality Comparison

Comparisons of the product quality are based on the metal composition of the PGM precipitates, recovered during thermal hydrolysis of the eluates, and the gold composition of the plated gold, recovered during electrowinning. It has been assumed, for the flowsheet alternatives producing multiple PGM precipitates, that only one overall precipitate was produced by combining the different products. This was based on the fact that a typical precious metal refinery operates on a sequential recovery bases, whereby each metal is recovered during a specific step in the overall process. It would thus be unlikely that refineries would treat multiple feed streams, as each metal can be recovered on its own in one process. It was thus assumed that only one feed is required for process in a PMR.

Figure 8.1 shows the PGM precipitate product compositions for the two CIS flowsheets, CIS 1 and CIS 2. It can be seen that the overall PGM grades of the products were high, at over 80%. These values compared favourably to average PGM grades of about 70% achieved during leaching of PGM automobile catalysts, as reported by Desmond *et al*. (1991), taking into account that other impurities were not tracked during simulation.

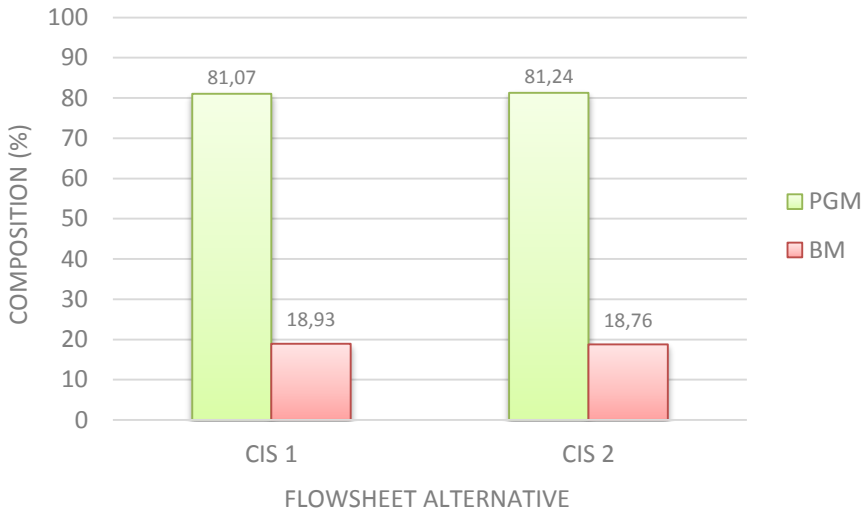

Figure 8.1: *PGM precipitate comparison (CIS flowsheets)*

In comparing the products of the different flowsheets to one another, it can be seen that the compositions were very similar, as the same general assumptions were made during simulation of the two flowsheets. Palladium made up the bulk of the products, at grades of over 52%. Platinum grades were around 27% while gold grades were only about 1.5%. The PGM grade of the CIS 2 precipitate was slightly higher than that of CIS 1 due to the increased recoveries observed from treating the high grade primary solution separately.

The observed compositions are, however, only relative to the other metals that were tracked during the simulations. In practise these precipitates would contain other gangue minerals such as silica, aluminium and various carbonates. The actual grade of these products would thus be lower than those presented in Figure 8.1. These products would be sufficient for the direct feed to a PMR if the gangue mineral content is lower than 15%, as typical PMRs can achieve optimum recoveries (> 99%) if the PGM feed grade is higher than 65% (Bezuidenhout, 2013). PMRs can, however, process feed streams with grades as low as 50% (Crundwell *et al*., 2011).

Figure 8.2 shows the gold product compositions of the two CIS flowsheets. Due to the sequential recovery methods employed very high grade products were produced, as the majority of the other metals precipitated during thermal hydrolysis, prior to electrowinning.

Figure 8.2: *Gold product compositions (CIS flowsheets)*

The gold content of the CIS 2 product was higher than that of the CIS 1 product, mainly due to the fact that the primary leach solution, containing all of the gold, was treated separately. The combination of the two leach solutions (CIS 1) resulted in a lower gold to base and precious metal ratio, thus resulting in a lower grade gold product. These products would still require further refining to produce a more marketable product.

8.2.2 Economic Comparisons

Economic comparisons between the different flowsheets were made by comparing the capital costs, operating costs and net present value. The net present value combines all of these variables and is the best parameter available to indicate the overall financial potential of a specific project.

8.2.2.1 Capital Costs

Figure 8.3 shows a comparison of the fixed capital costs per plant section between the two CIS flowsheets, CIS 1 and CIS 2. It was found that building two separate but smaller plants (CIS 2), as opposed to the single treatment option (CIS 1), required an increase in the capital investment of approximately 28%. This was as expected as it is well known that the cost of two components with the same total capacity as a single unit is more expensive when compare to one another (Peters & Timmerhaus, 1991; Turton *et al*., 2010). CIS 1 would thus be more favourable than CIS 2 if initial capital is limited.

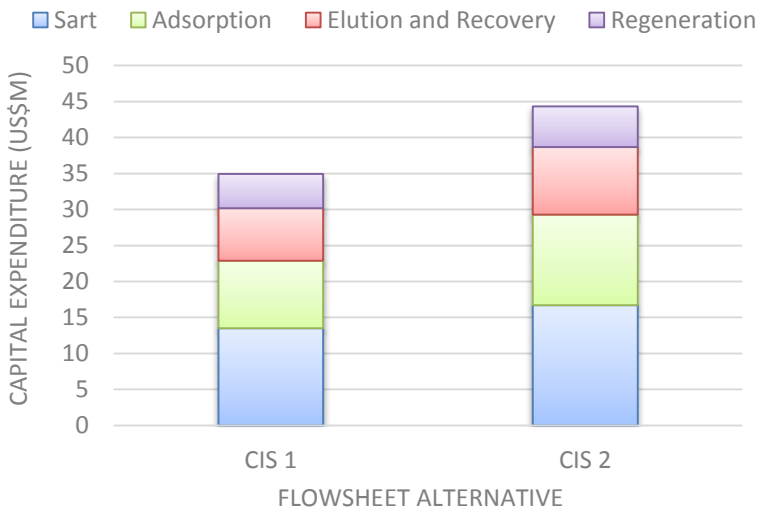

Figure 8.3: *Capital cost comparison (CIS Flowsheets)*

SART plant capital cost was found to be the major contributor to the total capital, at around 40% for both flowsheets. These costs were significant due to the large amounts of leach solution being treated. Improvements in the performance of the first stage bio-leach performed prior to the cyanide leach, to extract the base metals, could thus prove to be beneficial if this were to render a SART plant unnecessary. This would require almost complete extraction and is highly unlikely, based on current research (Mwase, 2009). This would be a less significant issue for a process leaching an ore with low base metal content.

Adsorption costs were estimated to amount to approximately 28% of the total capital costs. The costs saved by employing a pump cell configuration is significant if compared to the cost of the adsorption section of a plant utilizing a conventional adsorption circuit, which is typically 10 to 20% more expensive (Stange, 1999; Rogans & McArthur, 2002). As a result of the higher upgrade ratios that can be achieved in pump cell circuits, savings in elution capital are typically 25 to 50% (Stange, 1999; Rogans & McArthur, 2002). It can thus be seen that the adsorption section plays a critical role from an overall economic perspective. The rest of the costs were evenly split between the elution and recovery sections and carbon regeneration. The cost of the rotary kiln, although less pronounced when compared to the other sections (about 15%), was still a major contributor to the capital costs.

8.2.2.2 Operating Costs

Figure 8.4 shows a comparison of the operating costs per plant section between the two CIS flowsheets. It can be seen that the estimated operating cost of the two smaller plants (CIS 2) was significantly higher than that of the larger single plant (CIS 1). This was as expected as the labour costs and power requirements are rather insensitive to unit sizes but scale almost linearly as a result of the number of unit operations. Comparing these costs and noting that similar overall recoveries were achieved in both plants, the CIS 1 option would be preferred.

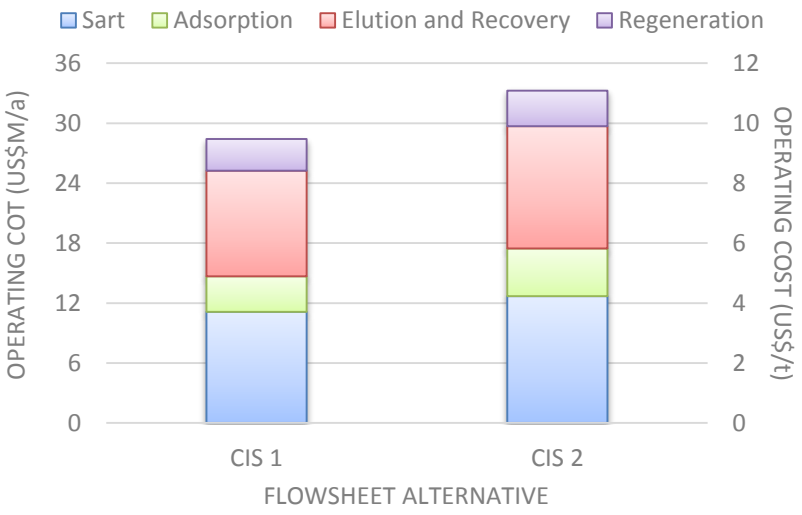

Figure 8.4: *Operating cost comparison (CIS flowsheets)*

The operating cost of the SART plant sections were determined to be the largest, at about 40%. The major costs associated with operating such a plant were the cost of the raw materials used to alter the pH of the solutions and the cost of sulphur addition ($NaHS$) to precipitate the base metals. SART plant labour costs were also the highest of all the plant sections.

Adsorption operating costs were low if compared to the SART and elution costs. No significant raw materials were required (carbon make-up costs were assigned to the regeneration circuits) while the major costs were the power and labour requirements, which were both low.

Elution and recovery operating costs were estimated to be high, at about 35%, due to the cost of the acid wash and pre-treatment solutions and the fact that the eluates were not used in multiple elution cycles. The major contributing costs were found to be the steam requirements for pre-heating of the acid and eluate and the power requirements of the autoclave to heat the solutions to 250°C.

Regeneration costs were found to be the lowest compared to the other plant sections. The high upgrade ratios achieved resulted in relatively low throughputs of carbon. Thus low operating costs were found, as fuel and steam requirements were low. Labour costs were also low as there was only one major process unit, the rotary kiln, in operation.

8.2.2.3 Potential Product Value

Figure 8.5 shows the estimated product values of the PGM and gold products for the two CIS flowsheets. The PGM product values were far superior to those of the gold products due to the low gold content of the ore. The estimated revenue of CIS 2 was higher, at US$ 268.1 million per annum as opposed to US$ 267.6 million per annum for CIS 1. This was due to the slightly larger and higher grade PGM precipitate produced. The larger gold value of CIS 1 value was due to the higher amount of base metals recovered due to electrowinning of the total solution and not only the primary solution. The amounts of gold recovered were, however, similar.

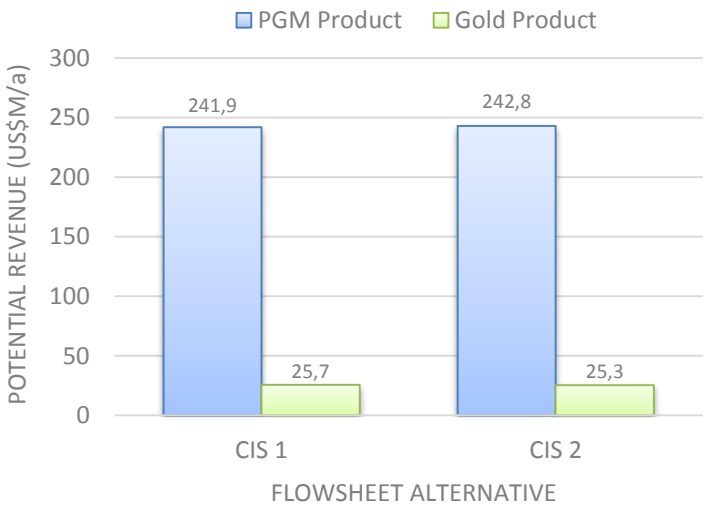

Figure 8.5: *Product value comparison (CIS flowsheets)*

8.2.2.4 Net Present Value

Figure 8.6 shows the estimated net present value of the two CIS flowsheets. It can be seen that the single treatment option's (CIS 1) project worth over a project life of 15 years was estimated to be about US$ 14 million more than that of the two plant option (CIS 2). This was mainly due to the significantly lower capital and operating costs of CIS 1 compared to CIS 2. Although CIS 2 had a slightly larger estimated yearly income, it could not cover the extra expenses required to achieve the desired performance.

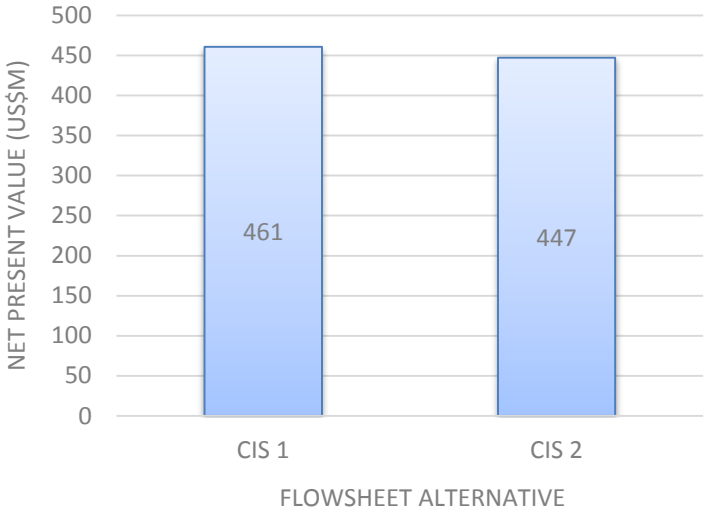

Figure 8.6: Net Present Value (NPV) comparison (CIS flowsheets)

8.2.2.5 Carbon-in-Solution Comparison Summary

From the preceding discussion the overall capital and operating costs were found to be lower for the single plant option (CIS 1) when compared to that of the separate plant option (CIS 2). CIS 2, however, produced a slightly larger, higher grade PGM product, but the increase in overall revenue due to this product could not cover the higher costs associated with constructing and operating two separate processing plants.

By taking all of the above results and comparisons into consideration, the single treatment option (CIS 1) should be considered if the need for a carbon-in-solution plant is pursued. The benefits of reduced capital and operating cost and a higher overall project profitability strongly outweighs the slightly smaller, lower product grades when compared to that of CIS 2.

8.3 Resin Flowsheet Comparison

The following section is a comparison between the five resin (RIS) flowsheet alternatives. The same general trends were observed for the separate plant options compared to the single plant options (RIS 1 vs. RIS 2 and RIS 3 vs. RIS 4). Discussions thus mainly focus on comparing the three options, RIS 1, RIS 3 and RIS 5 to one another.

8.3.1 Product Quality Comparison

Figure 8.7 shows the PGM precipitate product compositions for the five RIS flowsheets. It can be seen that the PGM grades were very high, at over 91%. This was due to the more selective adsorption nature of the resins for the precious metal over the base metals, and the fact that the resin acid wash can remove almost all nickel and zinc adsorbed.

Palladium made up the largest part of the products at close to 60%, while platinum was in excess of 30% for all cases. Gold grades were found to differ slightly (1.3% - 1.7%) as a result of the different gold recoveries observed between flowsheets. The flowsheets utilizing the elution with acidic thiourea (RIS 3, 4, 5) and/or the Minix resin (RIS 5) had slightly higher gold content.

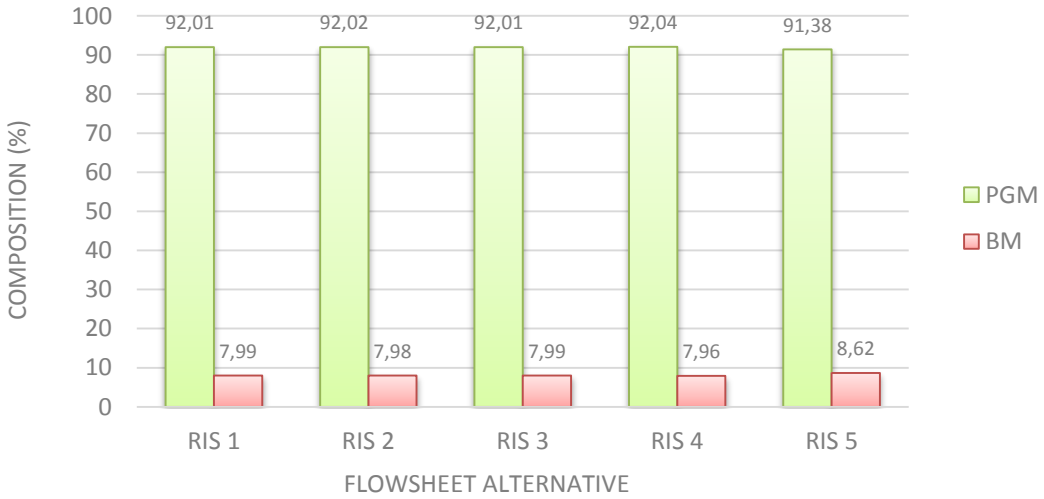

Figure 8.7: *PGM precipitate comparison (RIS flowsheets)*

In comparing the different products to one another there was very little difference between them, due to the similar assumptions made for all flowsheets. It should again be noted that these compositions would be lower during real plant operation as gangue minerals and other inert components were not tracked throughout the simulations. The amount of precipitate produced was also very similar for each flowsheet alternative, making it almost impossible to identify the process option that performed the best.

Figure 8.8 shows the gold product compositions for the five RIS flowsheets. It can be seen that the gold grades were very high, at over 91.5%. The high gold content of the last three flowsheet options were as a result of higher adsorption recoveries, elution efficiencies obtained with acidic thiourea and the higher gold selectivity of the Minix resin, in the case of RIS 5. The very high gold content of the RIS 5 product was due to the very low palladium, platinum and base metal recoveries obtained by the Minix resin and the fact that those metals that did adsorb were almost completely recovered during the preceding thermal hydrolysis step.

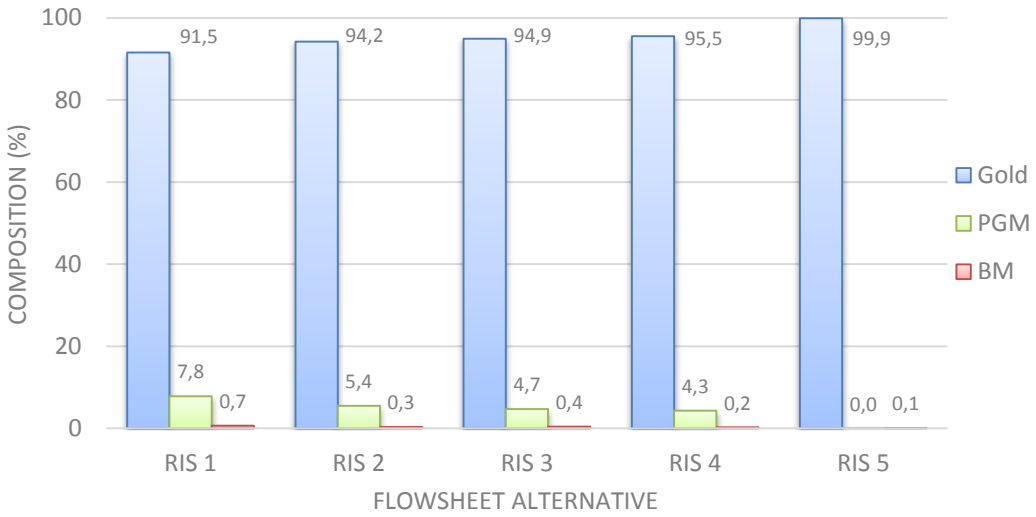

Figure 8.8: *Gold product comparison (RIS flowsheets)*

In comparison it can be seen that, with regards to the total precious metal grades of both products (PGM precipitate and the plated gold), the later three flowsheets options produced the best results.

8.3.2 Economic Comparisons

8.3.2.1 Capital Costs

Figure 8.9 shows a comparison of the fixed capital costs per plant section between the five RIS flowsheet options. Again it can be seen that the estimated costs of building two separate but smaller plants (RIS 2, 4) were higher than that of the single processing plant options (RIS 1, 3).

Capital costs for erecting the SART plants were found to be almost 60% of the total capital investment. If the need for a SART plant can be negated, significant cost savings could be achieved, as was mentioned earlier. Total SART costs were the same as those of the CIS flowsheets. The reason for the higher contribution of these costs in the RIS flowsheets is due to the costs saving of the other sections, compared to those of the CIS options.

Adsorption capital was found to be the second highest, at about 25% of the total capital investment. Although the resin adsorption tanks were small, the associated costs were still significant, and once again imply that efficient designs are required to ensure that overall capital costs are minimized. The adsorption capital costs were found to be similar for the single plant options (RIS 1, RIS 3), as well as for the two plant options (RIS 2, RIS 4). The adsorption capital of RIS 5 was, however, higher due to the fact that the secondary adsorption circuit was designed to treat the total capacity of leach solution.

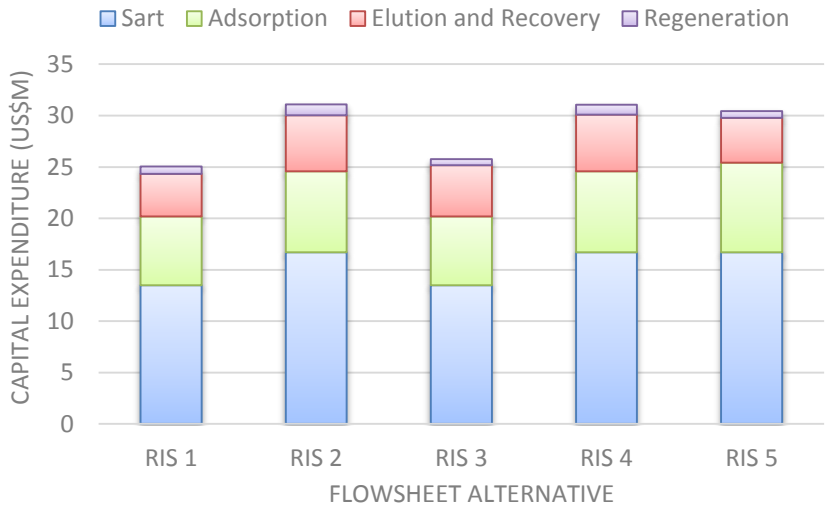

Figure 8.9: *Capital cost comparison (RIS flowsheets)*

Elution and recovery capital were low as a result of the high upgrade ratios achieved in the adsorption sections. Lower resin throughputs and the less severe operating conditions when compared to that used in the elution of carbon, resulted in major cost savings.

The estimated costs of the autoclaves were high due to the large volume of eluate required (15 BV) to achieve effective elution. This value was estimated from laboratory test work (Schoeman, 2012) and is very high if compared to that employed in industrial gold resin operations, which is usually 6 - 10 bed volumes (Kotze *et al*., 2005). Pilot plant testing should thus be performed to determine whether the elution of palladium and platinum could also be as effective at lower eluate requirements. This would result in considerable cost savings.

Regeneration costs were found to amount to less than 4% for all process alternatives. These low costs were due to the fact that no thermal regeneration circuits were required, and the assumption that the chemical regeneration of Amberlite PWA 5 eluted with zinc cyanide was to be performed in the elution columns. The only costs associated with regeneration were the raw material feed/storage tanks.

In comparing the flowsheets to one another, it would have been expected that the capital expenditure of the RIS 3 and RIS 4 options would be higher than that of RIS 1 and RIS 2, due to an extra recovery section required as a result of the sequential elution techniques employed. This was, however, not the case as the optimum loadings were found to be higher, resulting in less resin transferred to the elution sections, ultimately reducing the amount of eluate transferred to the subsequent recovery sections. The higher upgrade ratios, and the associated cost savings, countered the expected increase in capital costs. Although higher upgrade ratios increased resin inventory, fines, and value material locked-up, the higher gold recoveries, as a result of the more effective thiourea elution, resulted in these favourable trade-offs.

The capital costs of the more complex option (RIS 5) were found to be high. This was mainly due to the size of the secondary processing plant that had the same overall capacity as that of the overall RIS 1 and RIS 3 options. This was due to the fact that the platinum and palladium were not recovered in the primary adsorption circuit, and the resulting solution leaving that circuit had to be combined with the secondary leach solution.

Based on these capital costs, the RIS 1 or RIS 3 options are proposed if a resin-in-solution processing plant is required and limited capital is available.

8.3.2.2 Operating Costs

Figure 8.10 shows a comparison of the operating costs per plant section between the five RIS flowsheet options. Again it was estimated that the operating costs of the separate processing options would be much higher than that of the more simplistic single plant options.

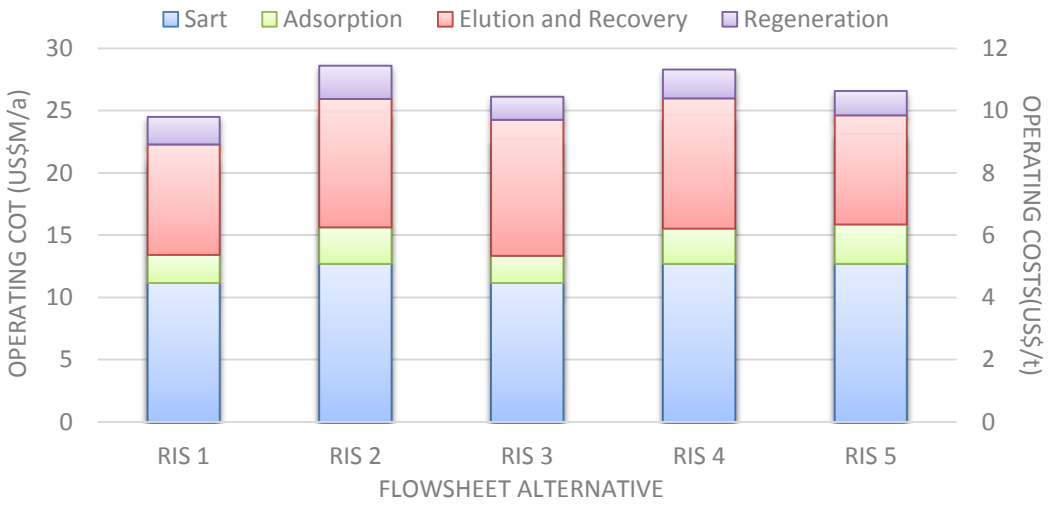

Figure 8.10: *Operating cost comparison (RIS flowsheets)*

The costs associated with the adsorption sections were low in comparison to the rest. Power requirements were low as a result of the small and relatively few adsorption tanks required, due to the fast adsorption kinetics of resins. Labour requirements were also low as the pump cell adsorption circuits are almost completely automated, while no raw materials were required during adsorption operation. Adjustments of the pH may be required but was deemed to be included in the contingency allowance. These costs could, however, be determined with the use of pilot plant test work.

SART and elution operating costs were estimated to be the main contributors, at about 40% and 35% of the total operating costs respectively. This high elution operating costs were mainly due to the fact that chemical reagents were used in stripping the metals from the resins. The cost of these chemicals are high and with the assumption that the eluates cannot be recycled back to the adsorption circuits, due to the expected degradation thereof during thermal hydrolysis, the replacement in each elution cycle proved to be significant.

This was less pronounced in the case of the elution with zinc cyanide, as most of the zinc cyanide could be recovered by regeneration of the resins. The costs associated with the elution with acidic thiourea were on the other hand high, as it was assumed that the thiourea complex would be destroyed. Thiourea is furthermore more expensive than zinc cyanide resulting in higher elution costs.

The lower elution costs using zinc cyanide as the eluting agent were offset by its associated regeneration costs. Thermal regeneration is not required but chemical regeneration of the adsorbed zinc cyanide needed to be performed. All resin-based flowsheets employed the zinc cyanide method as either the primary eluting agent or in combination with the thiourea method. The flowsheets that also used acidic thiourea (RIS 3, 4 and 5) did show lower estimated regeneration costs as resins eluted with thiourea do not require any chemical or thermal regeneration.

The operating cost of RIS 5 was found to be high, but lower than that of RIS 4. The sequential elution method used in RIS 4 was not required, as elution of the Minix resin only required one elution cycle, performed with acidic thiourea. The savings in the elimination of a sequential zinc cyanide elution and regeneration cycle can clearly be seen.

8.3.2.3 Potential Product Value

Product value comparisons for the different resin-in-solution flowsheets are presented in Figure 8.11. As the PGM recoveries of all flowsheets were similar, the associated income of these products is almost the same. The gold product value did vary between flowsheets as a result of the different adsorption capabilities of the two resins, and the effect that the elution efficiencies had on adsorption.

Lower gold product values were observed for the two flowsheets (RIS 1, 2) using Amberlite PWA 5 as the adsorbent and zinc cyanide as the eluting agent. The resins slower kinetics for gold adsorption, in comparison to those of platinum and palladium, and the high recycled loadings due to the ineffective elution of gold with zinc cyanide, resulted in low recoveries. The gold recovery, and hence the overall product value, were higher for the separate plant option (RIS 2), due to the higher gold concentration in the primary solution. It would prove to be very beneficial if zinc cyanide elution efficiencies of gold adsorbed onto Amberlite PWA 5 can achieve similar values as that achieved from other industrial strong base resins (>99%) (Fleming & Cromberge, 1984*a*, 1984*c*). Higher adsorption efficiencies would be achieved that in turn would result in significantly higher revenues.

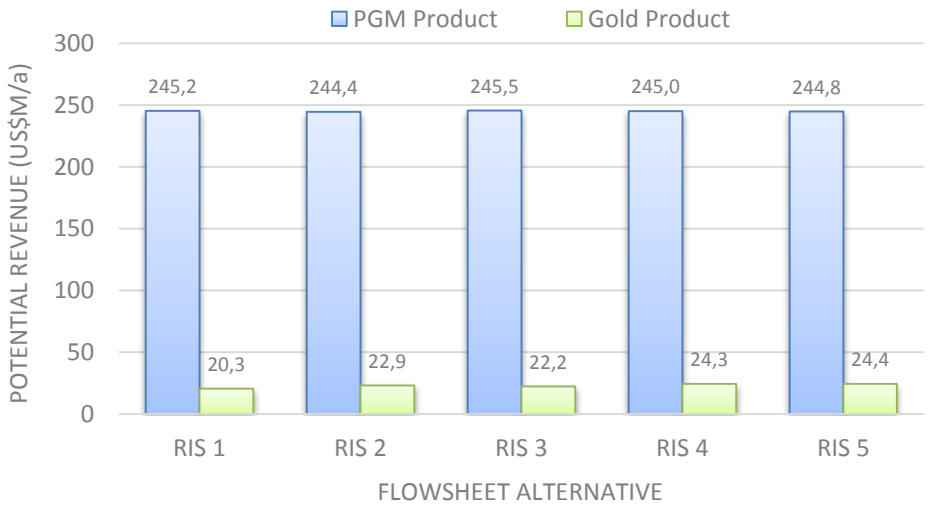

Figure 8.11: *Product value comparison (RIS flowsheets)*

It can further be seen that the gold product value increased for the flowsheet options employing acidic thiourea as the primary gold eluting agent (RIS 3, 4, 5). The lower recycled loadings, as a result of the more effective elution by thiourea, allowed for higher driving forces behind the adsorption mechanism, increasing overall adsorption. The same general increase in product value, as seen between RIS 1 and RIS 2, was observed between RIS 3 and RIS 4.

The gold product value for the RIS 5 flowsheet employing the Minix resin was only slightly higher than that of RIS 4, although significantly higher gold recoveries were achieved, as the high selectivity of the Minix resin resulted in a more pure gold product with very low PGM and base metal content, thus reducing the overall amount of product produced. It would be expected that the value of such a stream is higher but as it is difficult in assigning values to streams containing multiple metals, the effect thereof was not investigated.

8.3.2.4 Net Present Value

The estimated net present values, over a project life of 15 years, for the different RIS flowsheets are presented in Figure 8.12. As a result of the lower capital and operating costs of the RIS 1 and RIS 3 flowsheet options, and the fact that the overall product values of all flowsheets were similar, it was found that these two options would be the most profitable over the course of the project. The slightly increased overall recoveries of the three separate plant options (RIS 2, 4 & 5) could, however, not completely cover the increase in capital and operating costs.

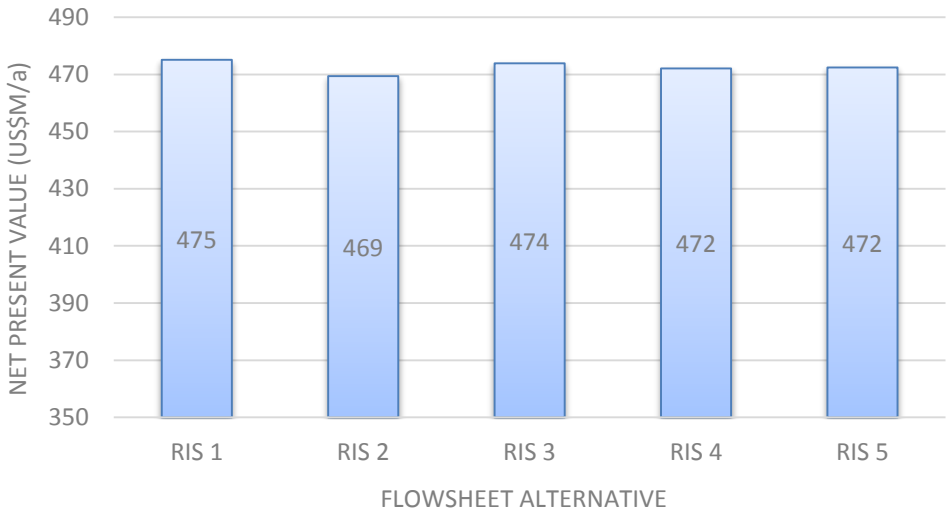

Figure 8.12: *Net Present Value (NPV) comparison (RIS flowsheets)*

8.3.2.5 Resin-in-Solution Comparison Summary

From the above discussions it was found that the overall capital and operating costs would be lower for the single plant options (RIS 1, 3) as opposed to those of the multiple plant options (RIS2, 4 & 5).The multiple plant options did, however, produce slightly higher grade products due to the increased recoveries, but the gain in overall revenue could not cover the higher costs associated with constructing and operating two separate processing plants.

By taking all of the above results and comparisons into consideration, the single treatment options (RIS 1, 3) should be considered if the need for a resin-in-solution plant is pursued. The benefits of reduced capital and operating costs and a higher overall project profitability, outweighs the slightly smaller, lower product grades if compared to that of the multiple plant options (RIS 2, 4 & 5).

The RIS 1 option would, however, be the clear favourite if the zinc cyanide elution efficiency of gold, adsorbed onto Amberlite PWA 5, could be similar to that achieved from other industrial resins eluted with zinc cyanide. This would allow for lower recycled loadings that would increase the driving force behind the adsorption mechanisms, greatly increasing overall gold recovery. The elution of gold with the use of acidic thiourea (RIS 3) was already high and additional benefits in increasing its respective elution efficiencies would not be as significant.

8.4 Overall Flowsheet Comparison

The following section is a comparison between the optimum carbon (CIS 1) and resin (RIS 1 & RIS 3) flowsheets, as identified in the previous sections. Based on the comparisons between these two different technologies, the best process option is proposed.

8.4.1 Product Quality Comparison

A comparison between the metal grades of the respective PGM precipitate products of the three flowsheet options (CIS 1, RIS 1 & RIS 3) are presented in Figure 8.13.

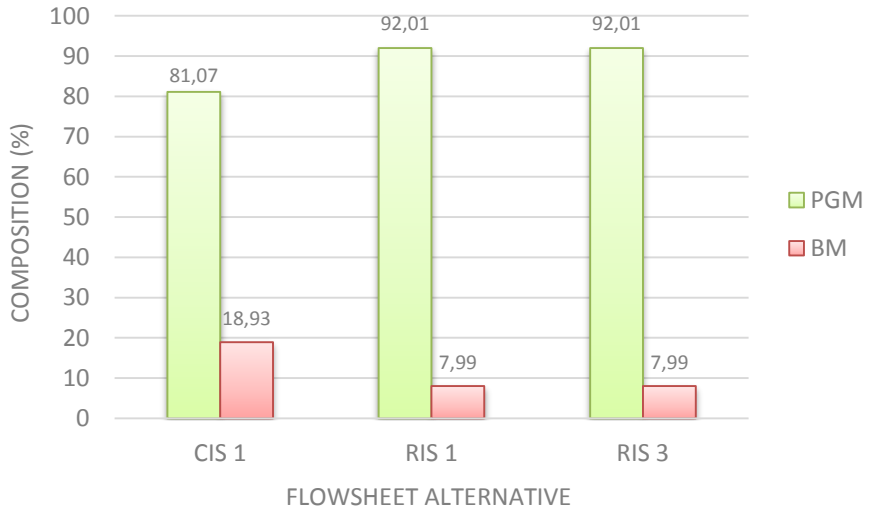

Figure 8.13: *PGM Precipitate product grade comparison (CIS vs. RIS)*

PGM grades of the RIS products were significantly higher than that estimated for CIS 1. The higher recoveries of the PGMs, lower recoveries of the base metals and the more effective acid washing achieved in the resin circuits, resulted in far superior product grades. The precipitates produced by the RIS flowsheets were furthermore smaller than that of the CIS 1 flowsheet, and subsequent savings in the handling, refining and smelting of these products could be expected.

The respective gold product grades are presented in Figure 8.14. The low gold recovery observed in the RIS 1 flowsheet resulted in a lower gold grade when compared to the other flowsheets, while the increased recovery of RIS 3, mainly due to the elution with thiourea followed by zinc cyanide, resulted in the product with the highest gold grade produced.

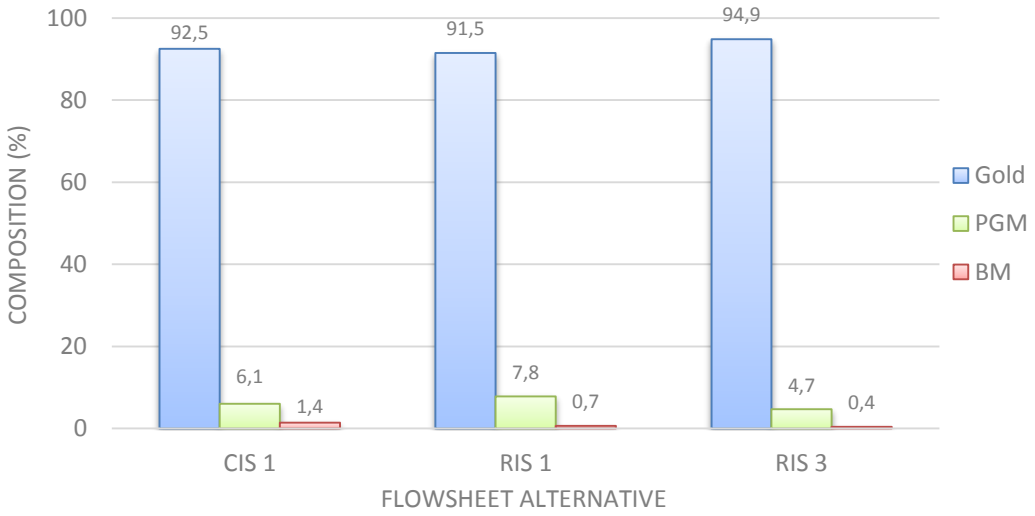

Figure 8.14: *Gold product grade comparison (CIS vs. RIS)*

The increased PGM recovery of the Amberlite PWA 5 resin and subsequent elution with zinc cyanide produced a higher PGM grade for RIS 1. The PGM grade of the RIS 3 gold product was lower due to the ineffective elution of platinum by acid thiourea. The precipitates produced by the RIS flowsheets were also smaller than that produced by CIS 1, and would result in subsequent savings in refining and smelting charges.

> *The RIS flowsheets thus produced higher grade products which would possibly result in the refinement thereof to be less challenging. These products were also smaller in total volume, and cost savings in handling of these products and the final refinement and smelting processes could be expected. Based on these factors, the RIS flowsheets would be preferred to the CIS 1 flowsheet.*

8.4.2 Economic Comparisons

8.4.2.1 Capital Costs

Figure 8.15 shows the capital cost comparison between CIS 1 and the two RIS flowsheets (RIS 1 & RIS 3). Overall capital cost savings of the RIS flowsheets over the CIS 1 flowsheet were found to be ±28%. This was lower than that reported by van Deventer *et al*. (2000) and Kotze *et al*. (2005) in the comparison between CIS and RIS for gold recovery, as SART plants were not included in their designs and economic evaluations.

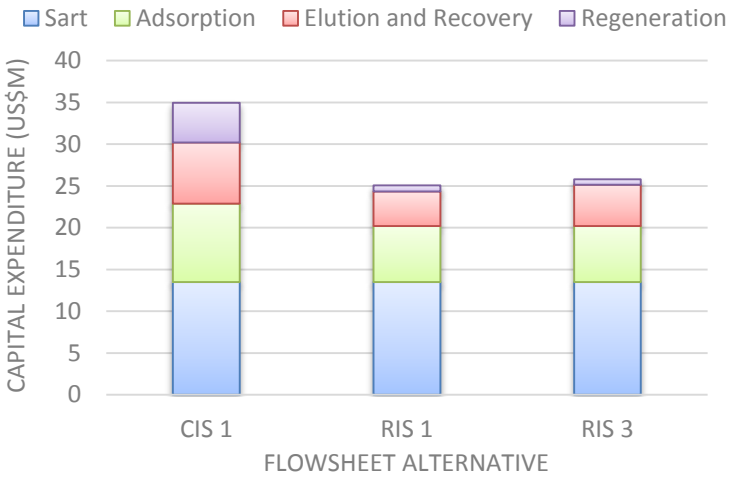

***Figure 8.15**: Capital cost comparison (CIS vs. RIS)*

SART capital costs were similar for all flowsheet alternatives, as the same processing plants were required. The RIS adsorption capital costs were about 35% lower than that of the CIS flowsheet, mainly due to the smaller sizes of the adsorption tanks required as a result of the improved adsorption kinetics of the resins.

The higher upgrade ratios achieved during resin adsorption had a big influence on the subsequent elution and recovery capital costs. The higher loadings resulted in less resin transferred to elution, requiring smaller acid wash and elution columns. Less eluates were also required that had a significant impact on the size of the autoclaves. The elution and recovery capital costs of the RIS flowsheets were thus almost 40% lower than that of the CIS 1 flowsheet.

The major cost saving between the RIS and CIS flowsheets were found to be the regeneration costs, mainly due to the fact that resins do not require thermal re-activation. The cost of the rotary kiln has proven to be significant in the case of the carbon-in-solution flowsheet. The resins only required, in the case of elution with zinc cyanide, a chemical regeneration step, and as this was assumed to be performed in the elution column, the costs thereof were minimal.

The RIS flowsheets thus showed decreased overall capital costs when compared to CIS 1 with the capital cost of the RIS 1 option being the lowest, 28.5% lower than that of CIS 1.

8.4.2.2 Operating Costs

Figure 8.16 shows the operating cost comparisons between the CIS and RIS flowsheets. Overall operating cost savings of the RIS flowsheets over the CIS flowsheet was found to be ±10%. This was lower than that reported by Kotze *et al*. (2005), mainly due the assumption that the eluting agents could not be recycled due to the expected degradation thereof during thermal hydrolysis. Operating costs are further also dependant on the ore treated and differences would be seen between different projects.

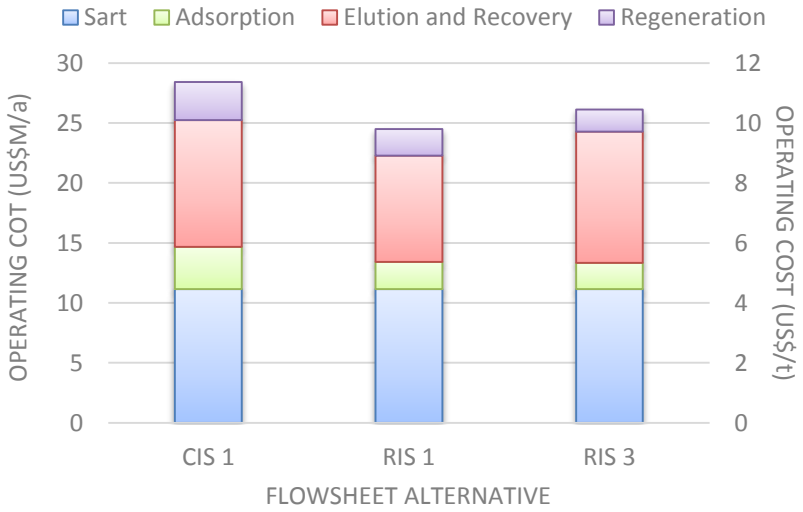

Figure 8.16: *Operating cost comparison (CIS vs. RIS)*

The adsorption operating cost of CIS 1 was higher than the RIS adsorption costs due to the larger adsorption tanks used. Power and utility requirements were thus estimated to be higher.

Elution costs were similar, with the RIS elution costs being slightly lower. Although smaller quantities of eluate were required for the RIS options, the cost of the chemical reagents proved significant. The effect thereof was more pronounced with regards to the elution with acidic thiourea (RIS 3), as the cost thereof is very high compared to de-ionized water, used for carbon elution, and zinc cyanide used primarily in the RIS 1 flowsheet. RIS 3 further employed three elution cycles and also required an extra thermal hydrolysis step due to the sequential elution method employed.

The regeneration costs of the RIS flowsheets were lower, as thermal regeneration was not required. The power, labour and utility requirements of the thermal regeneration of carbon were high and resulted in the subsequent high operating costs.

The RIS flowsheets thus showed decreased overall operating costs when compared to CIS 1 with the operating cost of the RIS 1 option being the lowest, 12.5% lower than that of CIS 1.

8.4.2.3 Potential Product Value

Product value comparisons for the different resin-in-solution flowsheets are presented in Figure 8.17. The PGM product values of the RIS products were found to be higher than that of the CIS 1 product, due to the increased recoveries estimated in the resin adsorption circuits and the fact that the resin circuits generated fewer fines. As a result thereof less precious materials were lost during resin screening.

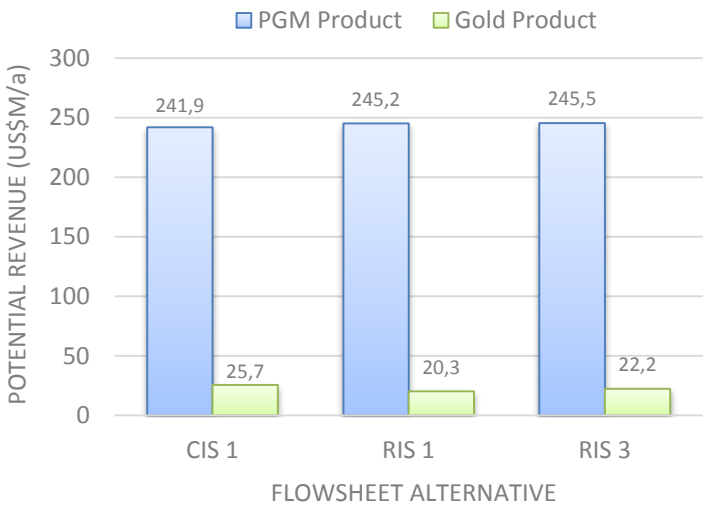

***Figure 8.17**: Product value comparison (CIS vs. RIS)*

The gold product value of the CIS 1 flowsheet option was estimated to be significantly higher than that of the RIS flowsheets. The gold recoveries of the RIS flowsheets were low, as a result of the more selective adsorption of platinum and palladium by the Amberlite PWA 5 resin. The gold product value of the RIS 3 option was higher than that of RIS 1 due to the higher recovery of gold due to the more effective elution with acidic thiourea.

The CIS 1 flowsheet option was thus estimated to generate total product value of US$ 267.6 million per annum while the RIS flowsheets were estimated at US$ 265.5 million per annum and US$ 267.7 million per annum respectively.

8.4.3 Profitability Comparisons

The overall selection between the three flowsheet options (CIS 1, RIS 1 & RIS 3) was made based on the net present value over a project life of 15 years. Monte Carlo simulations were also performed on the probability of these flowsheets to achieve these net present values, based on fluctuations in the capital costs, operating costs and the potential revenue.

8.4.3.1 Net Present Value

The estimated net present values (NPVs) of the three different flowsheets are presented in Figure 8.18. The estimated NPV of the CIS 1 flowsheet was the lowest. This was as a result of the higher capital and operating costs of this option compared to that of the RIS options.

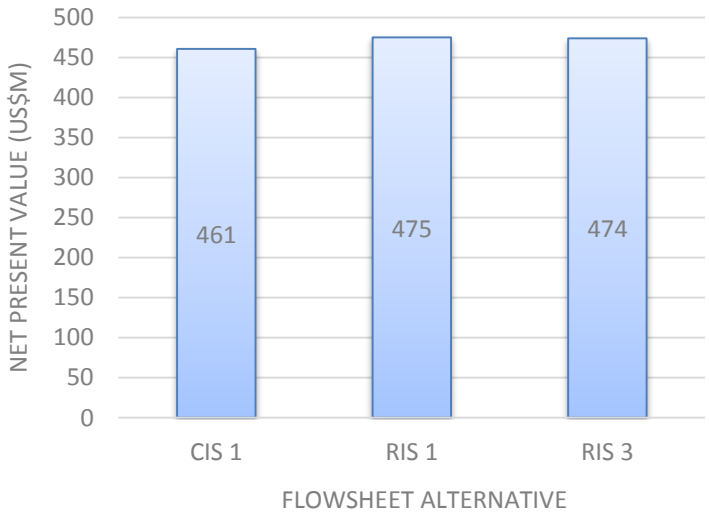

Figure 8.18: *Net Present Value (NPV) comparison (CIS vs. RIS)*

Although the potential product value of the RIS 1 flowsheet were estimated to be the lowest, capital and operating costs of this option were significantly lower than the rest, resulting in the estimated NPV thereof to be the highest. The extra potential income of the RIS 3 option was countered by the slightly higher capital and operating costs as a result of the sequential elution method employed.

> *The estimated NPVs of the RIS flowsheets were thus higher than that of the CIS 1 flowsheet, at an increase in the project worth of about 3%. The increase in project worth of the RIS process options, over a project life of 15 years, compared to that of CIS 1 were, however, not that significant. This was mainly due to the lower gold extraction of the Amberlite PWA 5 resin and the resulting elution methods required to counter this fact.*

To investigate the potential of these processes for their application to an ore reserve with low or no gold content, simulations were performed in which the gold content of the leach solutions were set to zero. An increase in project value of 8% for the RIS 1 options over that of CIS 1 was estimated for this specific application.

> *It can thus be seen that if an ore reserve with a lower gold content is to be treated, the expected benefits of the RIS 1 option would be more pronounced. Conversely, the CIS 1 option would be better for a higher gold to PGM ratio of the ore. The RIS flowsheets would, however, be more favourable in treating all PGM and gold ores if the issue of the lower gold recoveries obtained with these flowsheets could be addressed. Further investigations into this matter are thus required.*

8.4.3.2 Present Value Ratio

The NPV of a project is, however, greatly influenced by the level of the fixed capital investment. Although the estimated net present values were similar, a financial indicator that combined the initial investment with the overall project value was also calculated. This indicator, the present value ratio (PVR), is a measure of the overall gain in the project value as a function of the fixed capital investment, or otherwise stated the return on investment.

The PVR for each flowsheet alternative was calculated using equation (8.1).

$$PVR = \frac{Present\ Value\ of\ All\ Positive\ Cash\ Flows}{Present\ Value\ of\ All\ Negative\ Cash\ Flows} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (8.1)$$

$$= 1 + \frac{NPV}{FCI}$$

The fixed capital investment of the specific process options included the fixed capital costs, working capital, carbon and resin inventory costs and the associated value material lock-up.

A value of the PVR of less than 1 would indicate a process that would be unprofitable, while a value higher than unity would indicate a profitable project. The economic analysis of these projects did, however, not include the capital and operating costs associated with mining, leaching and the final product refining. The calculated PVR values are thus only an indication of the most profitable flowsheet alternative, and do not indicate whether the overall project would be financially favourable or not.

The estimated present value ratios of the three different flowsheet alternatives, normalized to the CIS 1 value, are presented in Figure 8.19.

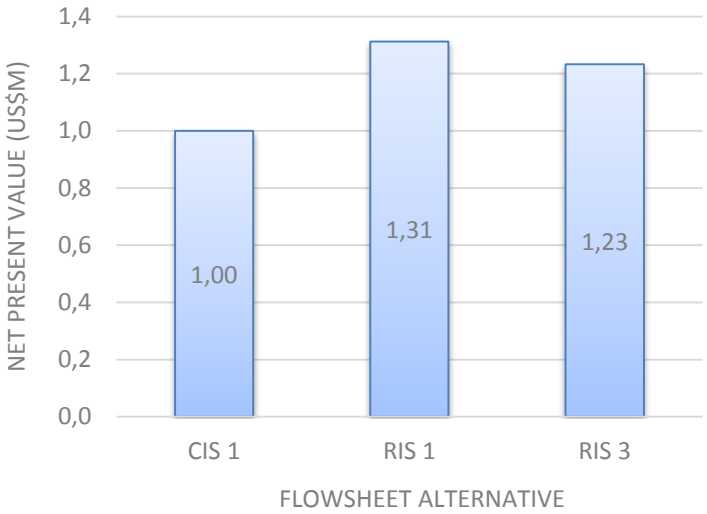

Figure 8.19: *Present value ratio (PVR) comparison (CIS vs. RIS)*

From Figure 8.19 it can be seen that the return on the initial capital investment, of the RIS flowsheets were superior to that of the CIS 1 option. The lower capital expenditure of the resin flowsheets proved to be a significant benefit. A 31% increase in project value over that of CIS 1 was estimated for the RIS 1 flowsheet option, while a 23% increase was estimated for the RIS 3 option.

RIS 1 operated at a slightly higher resin inventory than that of RIS 3, with the associated value mineral lock-up also being lower. The higher upgrade ratios required for RIS 3 to counter the increased capital and operating costs and to achieve the same net present value of RIS 1, resulted in higher loadings within the adsorption circuit that increased the value mineral lock-up, reducing the project worth.

> *Based on the present value ratio of these projects the RIS 1 flowsheet alternative proves to be the best available process option. The net worth of the project at the end of its lifetime would be the highest even though the lowest capital was required, leading to the return on investment being the highest, 31% higher than that of the CIS 1 flowsheet options..*

8.4.3.3 Monte Carlo Profitability Comparison

In order to assess the sensitivity of the profitability of the different flowsheet options to changes in the capital, operating and revenue costs, Monte Carlo simulations were performed. This simulation gives an indication as to the probability of the different flowsheets to achieve the estimated financial performance, i.e. the net present value. The upper and lower limits used for the revenue, capital costs and operating costs are given in Table 8.11. These values are commonly used in economic sensitivity analysis (Turton *et al*., 2010).

Table 8.11: *Upper and lower limits used for adjusting financial parameters*

	Revenue	*Operating Costs*	*Capital Costs*
Lower limit (%)	-20	-10	-20
Upper limit (%)	+5	+10	+30

In performing these simulations, one thousand different scenarios was created at various levels of the revenue, operating and capital costs, based on the upper and lower limits given in Table 8.11. The net present value of each of these scenarios was calculated where-after a cumulative distribution plot of the net present values was created.

The results of the Monte Carlo profitability simulations are presented in Figure 8.20 .

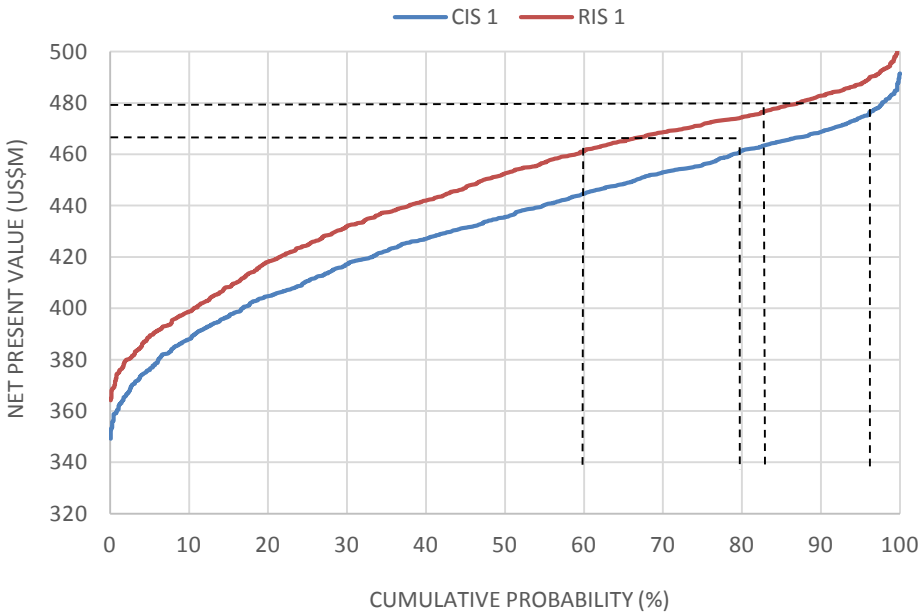

Figure 8.20: *Comparison of the NPV cumulative probability distribution*

From Figure 8.20 it can be seen that the CIS 1 flowsheet option has a predicted probability of about 20% to achieve the estimated NPV of US$ 461 million or higher over a project live of 15 years, based on the fluctuations in the different financial parameters as presented in Table 8.11. RIS 1 has, however, a slightly lower chance of achieving a higher NPV than that estimated for the base case, US$ 475 million, at about 17%. While there is a slightly higher risk in opting for the RIS 1 options over that of CIS 1, the predicted NPVs thereof are greater than that for CIS 1 for all situations simulated. In opting for the CIS 1 flowsheet, there is only a 4% chance that it would achieve the base case NPV of RIS 1, while there is an almost 40% chance that RIS 1 would outperform the base case of CIS 1.

> *Based on the results of the Monte Carlo sensitivity analysis it can be concluded that the best process option would be the RIS 1 flowsheet alternative, as there is a slightly higher financial security in utilizing this flowsheet option, to achieve a high NPV.*

8.4.3.4 Proposed Flowsheet Alternative

From the above discussions it can be seen that the RIS flowsheet options outperformed the CIS option in every aspect and it is proposed that such a process route be followed. The RIS 3 option showed promise for its application towards processing leach solutions obtained from heap leaching Platreef ore but there are several factors counting against it.

Firstly, elution is performed with a high cost acidic thiourea solution, which is known to decompose under the operating conditions required for effective elution. This eluting agent further results in ineffective elution of platinum from both the Amberlite PWA 5 and Minix resins which could be a major concern as it is not known at present whether the platinum is altered in such a way that the subsequent elution thereof would become problematic. Based on these factors the RIS 3 should not be considered.

The RIS 1 flowsheet alternative, however, proves to be the optimum process with no technical difficulties associated with it as all the proposed unit operations are used in the industry today. Although the overall precious metal recovery was slightly lower compared with the other flowsheets, it has the lowest capital investment and operating costs which led to the highest net present value with an excellent return on investment at a high probability of obtaining an economically favourable process. Based on these economic factors and overall performance this process is thus proposed for the recovery of precious metals from cyanide leach solutions.

Chapter 9

Conclusions and Recommendations

It this study it was found that the overall extraction and recovery of platinum group metals from low grade Platreef ore by cyanidation are possible and that several process options are available to accomplish this goal. Heap leaching could be used, that produces dilute leach solutions, with the platinum group metals being recovered from solution by either adsorption onto activated carbon or adsorption with ion exchange resins. These adsorbents can be eluted successfully producing concentrated solutions rich in PGMs that can be subjected to thermal hydrolysis and electrowinning to recover the PGMs in valuable, high-grade metallic precipitates, suitable for the direct feed to a PMR or to be toll refined.

Seven different flowsheet alternatives were developed based on these unit operations and variations thereof. With the use of data reported in literature, economic analyses and the use of kinetic adsorptions models for the simulation and optimization of these flowsheets, it was possible to arrive at an optimum design for each of the different flowsheet alternatives, based on maximizing the net present value.

It can be concluded that the overall performance of the resin-in-solution flowsheets were superior to that of the carbon-in-solution options, with regards to overall PGM recovery. The high selectivity of the Amberlite PWA 5 resin used in the simulations, for platinum and palladium proved to be a major contributor in achieving high PGM recoveries. Gold extraction was, however, lower in the RIS flowsheet options mainly due to the slow adsorption kinetics and low selectivity of this resin for gold and the fact that gold elution from this resin with zinc cyanide yield ineffective results. The gold concentrations in the feed streams to these processes were, however, significantly lower in comparison to those of platinum and palladium, and as a result the overall value mineral extraction efficiency of the resin flowsheets was higher.

The RIS flowsheets further produced higher grade products (>90% PGMs) than the CIS flowsheets (>80% PGMs) that could result in potential cost savings in refinement and smelting of the respective products.

From the cost analyses it was found that the capital and operating costs of the RIS flowsheets were lower than those of the CIS flowsheets, while the overall product value were similar. The superior adsorption kinetics of the resins proved to be a significant factor in that the adsorption circuits were smaller, higher metal loadings were achieved resulting in overall capital and operating cost savings due to smaller elution and regeneration circuits. This resulted in higher net present values over a project life of 15 years for the RIS flowsheets.

From all the comparisons made between the different flowsheet alternatives, the CIS 1 and RIS 1 options were identified as the best overall process options and are proposed if either of these methods/technologies were to be pursued for treatment of the Platreef ore. It was found that the RIS 1 option provides the largest return on investment with an increase in the overall return on investment of 31% over that of the CIS 1 option. There further proves to be a lower financial risk in choosing the RIS 1 flowsheet option as the NPV over a 15 year project life is superior to that of the CIS 1 option for large fluctuations in the revenue, operating and capital costs of the projects.

Based on these results it is proposed that a resin-in-solution flowsheet (RIS 1) be followed to process low grade Platreef ore, in which the ore is leached by means of a sequential bio- cyanide heap leach, with the PGM rich cyanide solution being treated in a SART plant, to remove base metals from solution in a sellable by product, while regenerating the associated cyanide for re-use in the leaching stages. The resulting solution should be transferred to a counter-current adsorption circuit employing the Amberlite PWA 5 resin, based on the AAC Pump Cell method. The loaded resin should then be eluted with zinc cyanide with the PGMs being recovered by thermal hydrolysis in an autoclave. This process was found to yield the most effective results with the lowest capital and operating costs required.

Whether resin-in-solution technology would, however, be preferred over that of activated carbon remains to be seen, and with the overall performance of the CIS 1 flowsheet not far behind that of RIS 1 and the fact that carbon adsorption is a well-known and well established process, it may be a more safer option to treat Platreef ore by means of an activated carbon process. This study has, however, shown that the potential for resin technology to provide more economically favourable results exists that could lead to the worldwide implementation thereof in place of activated carbon or froth flotation processes.

From this study several recommendations can be made as to improve the accuracy of the results obtained within this study to ultimately provide the mineral industry with sufficient information regarding the potential of these different flowsheet options to treat low grade PGM ores, and the Platreef resource in particular.

The literature and data available on process developments for the recovery of precious metals from cyanide leach solutions are extremely limited, with the studies that have been performed only at a concept/feasibility phase. Fundamental studies into the leaching of Platreef ore is required that could provide the necessary information to allow for more in-depth studies on recovering the platinum group metals from cyanide leach solutions. Furthermore, pilot plant studies should be performed that could be used in further process development, as studies has only been performed at laboratory scale.

Adding to these is the fact that limited sample material of Platreef ore was available for use in the experimental studies performed. Mainly synthetic leach solutions were thus used to replicate the expected leach solutions that would be obtained from leaching Platreef ore. The effects that impurities could have had on the experimental results were thus not determined.

The adsorption model employed furthermore does not take such effects into consideration. The optimum designs obtained in this study would thus possibly be a best case scenario and the effects of these impurities and other fouling material could reduce the actual operating performance of these units, especially adsorption and elution, considerably.

It is expected that the RIS designs would be the least affected as resins is not as susceptible to fouling as activated carbon although the kinetics of adsorption would be affected. It is thus proposed that the effect that impurities could have on the different processing units and their respective operation be quantified, as is currently the case at Stellenbosch University where extensive experimental test work is being performed with regards to carbon elution.

Furthermore there is currently no available kinetic adsorption models developed for the adsorption of platinum group metals onto activated carbon or ion exchange resins. The development of such models would prove to be beneficial in the design and optimization of flowsheet alternatives such as those provided in this project.

Although activated carbon and ion exchange resins used in the simulations have shown to be effective in extracting platinum, palladium and gold from cyanide solutions it is recommended that research be performed on developing an ion exchange resin specifically for the selective adsorption of these species, similar to the developmental work being performed at Mintek on the gold selective Minix resin.

By combining all of these, extensive work can be performed on developing a process for the extraction and recovery of platinum group metals from low grade PGM ores and dilute cyanide leach solutions which could prove valuable for the PGM industry.

References

Adams, M.D. (1991). Chemistry of adsorption and elution of gold on activated carbon. *Recent Developments in In-Pulp Technology*, October 7 - 8, The Southern African Institute of Mining and Metallurgy, Johannesburg, South Africa.

Adams, M.D. (1999). Aspects of the chemistry and mineralogy of gold-copper and copper-gold processing. *In: Adams, M.D. (Ed.), Processing of Gold-Copper and Copper-Gold Ores*. Oretest, Perth, pp. 17 - 42.

Adams, M.D., Friedl, J. and Wagner, F.E. (1995). The mechanism of adsorption of aurocyanide on to activated carbon, 2. Thermal stability of the adsorbed species. *Hydrometallurgy*, vol. 37, pp. 33 - 45.

Aguilar, M., Farran, A., and Marti, V. (1997). Capillary electrophoretic determination of cyanide leaching solutions from automobile catalytic converters. *Journal of Chromatography A*, vol. 778, pp. 397 - 402.

Allen, P., Lunt, D. and Messenger, P. (1998). Process selection, design and commissioning of the Rawas Gold Project. *In: TMS 1998 Conference*. The Minerals, Metals and Materials Society, USA.

Alibaba (2013). Alibaba.com Global Products. http://www.alibaba.com/ [Accessed 11 October 2013]

Bailey, P.R. (1987). Application of activated carbon to gold recovery. *In: Stanley, G.G. (Ed.), The Extractive Metallurgy of Gold in South Africa*, Monograph Series M7, vol. 1. The South African Institute of Mining and Metallurgy, Johannesburg, South Africa, pp. 379 - 614.

Barbetti, K., O'Leary, S. and Staunton, W. (2000). A Review of the AMIRA P420A gold industry survey. Gold Processing Technology P420A. Progress Report Module 2: Best Practice/ Benchmarking.

Barter, J., Lane, G., Mitchell, D., Kelson, R., Dunne, R., Trang, C. and Dreisinger, D. (2000). Cyanide management by SART. *In: TMS 2000 Conference.* The Minerals, Metals and Materials Society, USA.

References

Bezuidenhout, G. (2013). Personal Comminucation. Lonmin Plc, Rustenburg, South Africa.

Boshoff, P.T.E. (1994). Cyanide-free AARL elutions are feasible. *Minerals Engineering*, vol. 7, no. 2 - 3, pp. 251 - 264.

Botz, M., Kaczmarek, A. and Orser, S. (2011). Managing copper in leach solution at the çöpler gold project: laboratory testing and process design. Society of Mining, Metallurgy and Exploration Annual Meeting, Denver, Colorado, USA, February 27 - March 2.

Bryson, M. (2008). Review of Metallurgical Characteristics of Akanani Ore. *Mintek Restricted Presentation*, 22 January 2008.

Buson, G.D., Ngandu, D.S., le Roux, J.C. and Rogans, E.J. (1999). The West Driefontein reclamation carbon-in-pulp plant; pilot plant testwork, design, commissioning and optimization. *Journal of the South African Institute of Mining and Metallurgy*, vol. 99, no. 2, pp. 63 - 68.

Carrier, C.D., Hodouin, D. and Courchesne, M. (1987). Dynamic simulation of the CIP gold recovery process. *In: Proceedings of the CIM International Conference of Metallurgists - International Symposium on Gold Metallurgy*, Winnipeg, Montreal, pp. 309 - 325.

Chen, J. and Huang, K. (2006). A new technique for extraction of platinum group metals by pressure cyanidation. *Hydrometallurgy*, vol. 82, no 3 - 4, pp. 164 - 171.

CIBO (1997). *Energy Efficiency Handbook*. Council of Industrial Boiler Owners (CIBO), Burke, Virginia, USA.

Conradie, P., Johns, M. and Fowles, R. (1995). Elution and electrowinning of gold from gold-selective strong-base resins. *Hydrometallurgy*, vol. 37, pp. 349 - 366.

Cowley, A. (2013). Platinum 2013 Interim Review. Johnson Matthey Public Limited Company, Hertfordshire, London, England.

Cramer, L.A. (2001). The Extractive Metallurgy of South Africa's Platinum Ores. *Journal of the Minerals, Metals, and Materials Society*, vol. 53, no. 10, pp. 14 - 18.

Crundwell, F.K, Moats, M.S., Ramachandran, V., Robinson, T.G. and Davenport, W.G. (2011). *Extractive Metallurgy of Nickel, Cobalt and Platinum-Group Metals*. Elsevier, Amsterdam, The Netherlands.

Dai, X. and Breuer, P.L. (2009). Cyanide and copper cyanide recovery by activated carbon. *Minerals Engineering*, vol. 22, pp. 469 - 476.

Dai, X., Jeffrey, M. and Breuer, P.L. (2010). A mechanistic model of the equilibrium adsorption of copper cyanide species onto activated carbon. *Hydrometallurgy,* vol. 101, pp. 99 - 107.

Davidson, R.J. and Duncanson, D. (1977). The elution of gold from activated carbon using deionized water. *Journal of the South African Institute of Mining and Metallurgy*, vol. 77, no. 12, pp. 254 - 261.

Davidson, R.J. and Schoeman, N. (1991). The management of carbon in a high-tonnage CIP operation. *Journal of the South African Institute of Mining and Metallurgy*, vol. 91, no. 6, pp. 195 - 208.

Davidson, R.J. and Schmidt, C.G. (1986). A review of the AARL process for the elution of gold from activated carbon. *In: Proceedings of the 13th Congress of the Council of Mining and Metallurgical Institutions*, Singapore, pp. 215 - 225.

Davidson, R.J. and Veronese, V. (1979). Further studies on the elution of gold from activated carbon using water as the eluent. *Journal of the South African Institute of Mining and Metallurgy*, vol. 79, no. 15, pp. 437 - 445.

Desmond, D.P., Atkinson, G.B., Kuczynski, R.J., Walters, L.A. (1991). High-temperature cyanide leaching of platinum-group metals from automobile catalysts - laboratory tests, *Report of Investigation 9384*, United States Bureau of Mines.

Dionex Corporation (2003). Direct Determination of Metal Cyanides by Ion Chromatography with UV Absorbance Detection, *Application Update 147*, LPN 1570 PDF 1/04, 2003

Dippenaar, A. and Proudfoot, M.R. (2005). The performance of the AAC Pumpcell circuits at the Gold Fields Limited Driefontein and Kloof Operations. *Randol Perth Forum 2005*, Perth, Australia.

Dixon, S., Cho, E.H. and Pitt, C.H. (1978). The interaction between gold cyanide, silver cyanide and high surface area charcoal. *American Institute of Chemical Engineers Symposium Series*, vol. 74, no. 173, pp. 75 - 83.

Dreisinger, D., Molnar, R., Mezei, A., Baxter, K. and Wardell-Johnson, M. (2009). The recovery of nickel and cobalt from the Northmet deposit using the Platsol Process with production of mixed or separate nickel and cobalt hydroxides. *Proceedings of the 39t Annual Hydrometallurgy Meeting held in conjunction with the 48t Annual Conference of Metallurgists of CIM,* Sudbury, Ontario, Canada

Eksteen, J.J., Mwase, J.M., Peterson, J., Bradshaw, S.M., Mpinga, N. and Snyders, C.A. (2012). A novel energy efficient process for the extraction of platinum group metals through a sequential stage high temperature heap leach process and subsequent recovery and elution using activated carbon. *Proceedings of Biohydrometallurgy '12*, Falmouth, UK.

Estay, H., Becker, J., Carvajal, P. and Arrigada, F. (2012). Predicting HCN gas generation in the SART process. *Hydrometallurgy*, vol. 113 - 114, pp. 131 -142.

Fleming, C.A. (1992). Hydrometallurgy of precious metals recovery. *Hydrometallurgy*, vol. 30, pp. 127 - 162.

Fleming, C. (1993). CIP and RIP: Where to next? *In: Hisky, J. and Warren, G. (Eds.), Hydrometallurgy Fundamentals, Technology and Innovation*, pp. 379 - 393.

Fleming, C.A. (2002). CIP/CIL/CIC Adsorption circuit process selection. *In: Mineral Processing Design, Practice, and Control Proceedings*. The Minerals, Metals and Materials Society, USA.

Fleming, C. and Cromberge, G. (1984*a*). The elution of aurocyanide from strong-and weak-base resins. *Journal of the South African Institute of Mining and Metallurgy*, vol. 84, no. 9, pp. 269 - 280.

Fleming, C. and Cromberge, G. (1984*b*). The extraction of gold from cyanide solutions by strong- and weak-base anion exchange resins. *Journal of the South African Institute of Mining and Metallurgy*, vol. 84, no. 5, pp. 125 - 137.

Fleming, C. and Cromberge, G. (1984*c*). Small-scale pilot-plant tests on the resin-in-pulp extraction of gold from cyanide media. *Journal of the South African Institute of Mining and Metallurgy*, vol. 84, no. 11, pp. 369 - 378.

References

Fleming, C.A., Mezei, A., Bourricaudy, E., Canizares, M. and Ashbury, M. (2011). Factors influencing the rate of gold cyanide leaching and adsorption on activated carbon, and their impact on the design of CIL and CIP circuits. Minerals Engineering, vol. 24, pp. 484 - 494.

Fleming, C.A., Nicol, M.J. and Nicol, D.I. (1980). The optimization of a carbon-in-pulp adsorption circuit based on the kinetics of extraction of aurocyanide by activated carbon. *In: Proceedings of the Symposium on Ion Exchange and Solvent Extraction in Mineral Processing*, National Institute for Metallurgy, Randburg, South Africa.

Fisher, M.J. and LaBrooy, S.R. (1997). The effect of nickel on gold recovery during adsorption and elution. *World Gold '97 Conference*, Singapore, September 1 - 3.

Ford, K.J.R., Fleming, C.A and Henderson, R.D. (2008). Application of the SART process to heap leaching of gold-copper ores at Maricunga, Chile. *40th Annual Meeting of the Canadian Mineral Processors*, Ottawa, January 22 - 24.

Gain, S.B. and Mostert, A.B. (1982). The geological setting of the platinoid and base metal sulphide mineralization in the Platreef of the Bushveld Complex in Drenthe, North of Potgietersrus. *Economic Geology*, vol. 77, no. 6, pp. 1395 - 1404.

Gerrard, A.M. (2000). *Guide to capital cost estimating* (4th Edition). *Institution of Chemical Engineers*, UK.

Green, B.R., Smit, D.M.C., Maumela, H. and Coetzer, G. (2004). Leaching and recovery of platinum group metals from UG-2 concentrates. *Journal of the South African Institute of Mining and Metallurgy*, vol. 104, no. 6, pp. 323 - 332.

Greenhalgh, L.P.H., Riley, R.P., Blois, M.D.S., Talocchino, L. (1991). Cost benefits of Continuous Elution of gold from activated carbon. *Proceedings of the Randol Gold Forum '91,* Cairns, Australia, 1991, pp. 379 - 384.

Hampton, P.A. (2002). Zinc cementation - The Merrill Crowe process. *In: Mineral Processing Design, Practice, and Control Proceedings*. The Minerals, Metals and Materials Society, USA, pp. 1663 - 1679.

Heinen, H.J., Peterson, D.G. and Lindstrom, R.E. (1978). Processing gold ores using heap leach-carbon adsorption methods. *Information Circular, 8770*. United States Bureau of Mines.

References

InfoMine (2013). Comodity and metal prices. *InvestmentMine Mining Markets & Investment.* http://www.infomine.com/investment/metal-prices/ [Accessed 2 November, 2013].

Johns, M.W. (1986). Model application. *In: SAIMM Carbon School Design and Optimization of Plants for the Recovery of Gold by Activated Carbon*, Mintek, September, 1985, Lecture 5. South African Institute of Mining and Metallurgy, Johannesburg, South Africa.

Jollie, D. (2009). Platinum 2009 Interim Review. Johnson Matthey Public Limited Company, Hertfordshire, London, Engalnd.

Jones, M., Macrae, C., Sparrow, G. Woodcock, J.T. (1988). Distribution of inorganics and gold on the external and internal surfaces of activated carbon from CIP plants. *The Australian Institute of Mining and Metallurgy Bulletin and Proceedings*, vol. 293, no. 6, pp. 121 - 126.

Kappes, D.W. (2005). Heap leaching of gold and silver ores. *In: Adams, M.D. (Ed.), Developments in Mineral Processing*, Vol. 15, Elsevier, Amsterdam, The Netherlands, pp. 456 - 478.

Kemix (2013). Personal Communication. Kemix (Pty) Ltd. Midrand, Johannesburg, South Africa.

Kitney, M. (1998). Cyanide regeneration of gold tailings - Goldconda Beaconsfield experience. *In: Perth International Gold Conference*. Randol International Ltd, Colorado, USA.

Kotze, M., Green, B., Mackenzie, J. and Virnig, M. (2005). Resin-in-pulp and resin-in-solution. *In: Adams, M.D. (Ed.), Developments in Mineral Processing,* Vol. 15, Elsevier, Amsterdam, The Netherlands, pp. 603 - 635.

Kuczynski, R.J., Atkinson, G.B. and Walters, L.A. (1992). High-temperature cyanide leaching of automobile catalysts - process development unit. *Report of Investigation 9428*, United State Bureau of Mines.

La Brooy, S.R., Linge, H.G. and Walker, G.S. (1994). Review of gold extraction from ores. *Minerals Engineering*, vol. 7, no. 10, pp. 1213 - 1242.

La Brooy, S.R., Hosking, J.W., Muir, D.M., Ruane, M., Smith, I. and Hinchliffe, W.D. (1984). Studies on the fouling and carbon regeneration of CIP carbon. The Australian Institute of Mining and Metallurgy Perth and Kalgoorlie Branches, Regional Conference on "Gold-Mining, Metallurgy and Geology", pp. 257 - 269.

Laxen, P.A., Becker, S.M. and Rubin, R. (1979). Developments in the application of carbon-in-pulp to the recovery of gold from South African ores. *Journal of the South African Institute of Mining and Metallurgy*, vol. 79, no. 11, pp. 315 - 327.

Liddell, K.S. and Adams, M.D. (2012). Kell hydrometallurgical process for extraction of platinum group metals and base metals from flotation concentrates. *Journal of the South African Institute of Mining and Metallurgy*, vol. 112, no. 1, pp. 31 - 36.

Liebenberg, C.J. (2012). Recovery of base metals from a sulphate-based bioleach solution using commercially available chelating ion exchange resins and adsorbents. *MSc Thesis*, Department of Process Engineering, Stellenbosch University, South Africa.

Liebenberg, S.P. and van Deventer, J.S.J. (1998). A dynamic model of the carbon-in-pulp process incorporating shifting adsorption equilibria. *Minerals Engineering*, vol. 11, no. 6, pp. 551 - 562.

Lu, J., Dreisinger, D.B. and Cooper, W.C. (2002). Thermodynamics of the aqueous copper-cyanide system. *Hydrometallurgy*, vol. 66, no. 1 - 3, pp. 23 - 36.

Lunt, D. and Weeks, T. (2005). Process flowsheet selection. *In: Adams, M.D. (Ed.), Developments in Mineral Processing,* Vol. 15, Elsevier, Amsterdam, The Netherlands, pp. 73 - 96.

MacPhail, P.K., Fleming, C.A. and Sarbutt, K.W. (1998). Cyanide Recovery by the SART Process for the Lobo-Marte Project, Chile. *Randol Gold and Silver Forum*, Denver, Colorado, USA, April 26 - 29.

Marsden, J.O. and Fuerstenau, M.C. (1993). Comparison of Merrill Crowe precipitation and carbon adsorption for precious metals recovery. *Proceedings of* the *XVIII International Mineral Processing Congress*, Australian Institute of Mining and Metallurgy, Sydney, Australia, pp. 1189 - 1194.

Marsden, J.O. and House, C.I. (2006). *The Chemistry of Gold Extraction* (2nd Edition). Society of Mining, Metallurgy and Exploration, Littleton, Colorado, USA.

McInnes, C.M., Sparrow, G.J. and Woodcock, J.T. (1994). Extraction of platinum, palladium, and gold by cyanidation of Coronation Hill ore. *Hydrometallurgy*, vol. 35, pp. 141 - 159.

Milbourne, J., Tomlinson, M., Gormely, L., (2003). Use of hydrometallurgy in direct processing of base metal/PGM concentrates. *Hydrometallurgy 2003 - 5th International Conference in Honor of Professor Ian Ritchie - Volume 1: Leaching and Solution Purification*, The minerals, metals and materials society, pp. 617 - 630.

Mpinga, C.N. (2012). The extraction of precious metals from alkaline cyanided medium by granular activated carbon. *MSc Thesis*, Department of Process Engineering, Stellenbosch University, South Africa.

Mpinga, C.N., Bradshaw, S.M., Akdogan, G., Snyders, C.A. and Eksteen, J.J. (2013). Evaluation of the Merrill-Crowe process for the simultaneous removal of platinum, palladium and gold from cyanide leach solutions. *Hydrometallurgy*, Available online 11 November 2013: http://www.sciencedirect.com/science/article/pii/S0304386X13002223

Mwase, J.M. (2009). Hydrometallurgical extraction of platinum group metals from low-grade ore concentrate. *MSc Thesis*, University of Cape Town, South Africa

Mwase, J.M. and Peterson, J. (2012). High level cost estimation - Akanani. *Internal Report*, Department of Process Engineering, Stellenbosch University, South Africa.

Mwase, J.M., Petersen, J. and Eksteen, J.J. (2012*a*). A conceptual flowsheet for heap leaching platinum group metals (PGMs) from a low-grade ore concentrate. *Hydrometallurgy*, vol. 111 - 112, pp. 129 - 135.

Mwase, J.M., Petersen, J. and Eksteen, J.J. (2012*b*). Assessing a two-stage heap leaching process for Platreef flotation concentrate. *Hydrometallurgy*, vol. 129 - 130, pp. 74 - 81.

Mwase, J.M., Petersen, J. and Eksteen, J.J. (2014). A novel sequential heap leach process for treating crushed Platreef ore. *Hydrometallurgy*, vol. 141, pp. 97 - 104.

Mogosetsi, D. (2006). Scoping flotation work on sample from the Akanani platinum project. *External Mintek Restricted Report*, 4272.

Nicol, M.J., Fleming, C.A. and Cromberge, G. (1984*a*). The adsorption of gold cyanide onto activated carbon. I. The kinetics of adsorption from pulps. *Journal of the South African Institute of Mining and Metallurgy*, vol. 84, no. 2, pp. 50 - 54.

References

Nicol, M.J., Fleming, C.A. and Cromberge, G. (1984*b*). The adsorption of gold cyanide onto activated carbon. II. The kinetics of adsorption from pulps. *Journal of the South African Institute of Mining and Metallurgy*, vol. 84, no. 3, pp. 70 - 78.

Nicol, M.J., Fleming, C.A. and Cromberge, G. (1984*c*). The adsorption of gold cyanide onto activated carbon. III. Factors influencing the rate of loading and the equilibrium capacity. *Journal of the South African Institute of Mining and Metallurgy*, vol. 84, no. 4, pp. 85 - 93.

Peters, M.S. and Timmerhaus, K.D. (1991). *Plant design and economics for chemical engineers* (4th Edition). McGraw-Hill, New York, USA.

Peterson, J. (2012). Personal Communication. University of Cape Town, South Africa.

Renner, H., Schlamp, G., Drost, I., Lüschow, H. M., Tews, P., Panster, P., Diehl, M., Lang, J., Kreuzer, T. K. A., Starz, K. A., Dermann, K., Rothaut, J., Drieselmann, R., Peter, C. and Schiele, R. (2001). Platinum Group Metals and Compounds. *In: Ullmann's Encyclopedia of Industrial Chemistry*, Wiley-VCH, Weinheim, pp. 317 - 388.

Rogans, J. and McArthur, D. (2002). The evaluation of the AAC Pump-Cell circuits at Anglogold's West Wits operations. *Journal of the South African Institute of Mining and Metallurgy*, vol. 102, no. 4, pp. 181 - 188.

Roijals, O., Martf, V., Meinhardt, E., Cortina, J.L. and Aguilar, M. (1996). Characterization of spent automotive catalyst residues for precious metal recycling using hydrometallurgical technologies. *Chemical Industry and Environment*, vol. 2, pp. 419 - 428.

Ruane, M. (1982). Comparison of the Zadra, Anglo-American and organic procedures for desorption of gold from activated carbon. *The Australian Institute of Mining and Metallurgy Perth and Kalgoorlie Branches and Murdoch University, Carbon-In-Pulp Seminar*, pp. 393 - 414.

Seymour, R.J. and O'Farrelly, J.I. (2001). Platinum Group Metals, *Kirk - Othmer Encyclopedia of Chemical Technology - Online*. Published online: 13 July 2001 http://onlinelibrary.wiley.com/book/10.1002/0471238961

References

Sceresini, B.J.S. (1982). The design, construction and operation of a 500,000 tonnes per annum carbon-in-pulp plant at Kambalda, Western Australia. *The Australian Institute of Mining and Metallurgy Perth and Kalgoorlie Branches and Murdoch University, Carbon-In-Pulp Seminar*, pp. 237 - 278.

Sceresini, B.J.S. (2005). Gold-copper ores. *In: Adams, M.D. (Ed.), Developments in Mineral Processing*, Vol. 15, Elsevier, Amsterdam, The Netherlands, pp. 789 - 824.

Schoeman, E. (2012). Recovery of platinum, palladium and gold from a pregnant cyanide heap leach solution, with the use of ion exchange resins. *MSc Thesis*, Department of Process Engineering, Stellenbosch University, South Africa.

Schoeman, E., Bradshaw, S.M., Akdogan, G. and Eksteen, J.J. (2012). The recovery of platinum, palladium and gold from a cyanide heap leach solution, with the use of ion exchange resins. *Proceedings of The South African Institute of Mining and Metallurgy, Platinum 2012*, Johannesburg, South Africa

Schoeman, N., Rogans, E.J. and MacIntosh, A.J. (1996). AAC Pump-Cells, a cost effective means of Gold Recovery from slurries. *South African Institute of Mining and Metallurgy Hidden Wealth Conference*, Johannesburg, pp. 173 - 179.

Schouwstra, R.P. and Kinloch, E.D. (2000). A short geological review of the Bushveld Complex. *Platinum Metals Review*, vol. 44, no. 1, pp. 33 -39.

Shamaila, S. and O'Connor, C.T. (2008). The role of synthetic minerals in determining the relative flotation behaviour of Platreef PGE tellurides and arsenides. *Minerals Engineering*, vol. 21, pp. 899-904.

Smith, R. (2005). Chemical process design and integration (2nd Edition). Wiley and Sons Ltd, West Sussex, England.

Snyders, C.A. (2011). A fundamental study of PGM elution from activated carbon. *Internal Report*, Department of Process Engineering, Stellenbosch University, South Africa.

Snyders, C.A., Mpinga, C.N., Bradshaw, S.M., Akdogan, G. and Eksteen, J.J. (2013). The application of activated carbon for the adsorption and elution of platinum group metals from dilute cyanide leach solutions. *Journal of the South African Institute of Mining and Metallurgy*, vol. 113, no. 5, pp. 381 - 388.

Stange, W. (1991). The optimization of the CIP process using mathematical and economic models. *Minerals Engineering*, vol. 4, no. 12, pp. 1279 - 1295.

Stange, W. (1999). The process design of gold leaching and carbon-in-pulp circuits. *Journal of the South African Institute of Mining and Metallurgy*, vol. 99, no. 1, pp. 13 - 26.

Stange, W. and King, R.P. (1987). A population balance approach to the modelling of the CIP process. *In: King, R.P. and Barker, I.J. (Eds.), Proceedings of APCOM87*, The 20th International Symposium on the Application of Computers and Mathematics in the Minerals Industries, vol. 2, Johannesburg, South Africa, pp. 209 - 221.

Stange, W., King, R.P. and Woollacott, L.C. (1990*a*). Towards more effective simulation of CIP and CIL processes. 2. A population-balance-based simulation approach. *Journal of the South African Institute of Mining and Metallurgy*, vol. 90, no. 11, pp. 307 - 314.

Stange, W., Woollacott, L.C. and King, R.P. (1990*b*). Towards more effective simulation of CIP and CIL processes. 3. Validation and use of a new simulator. *Journal of the South African Institute of Mining and Metallurgy*, vol. 90, no. 12, pp. 323 - 331.

Staunton, W.P. (2005). Carbon-in-pulp. *In: Adams, M.D. (Ed.), Developments in Mineral Processing,* Vol. 15, Elsevier, Amsterdam, The Netherlands, pp. 73 - 96.

Stewart, M. and Kappes, D. (2012). SART for copper control in cyanide heap leaching. *Journal of the South African Institute of Mining and Metallurgy*, vol. 112, no. 12, pp. 1037 - 1043.

Steyn, J. and Sandenbergh, R.F. (2004). A study of the influence of copper on the gold electrowinning process. *Journal of the South African Institute of Mining and Metallurgy*, vol. 104, no. 3, pp. 177 - 182.

Torres, V.M. and Costa, R.S. (1997). Recovery and production of gold and platinum-group metals by using cyanidation under pressure, *Braindex online database*.

Tran, T., Nguyen, H.H., Hsu, Y.J. and Wong, P.L.M. (1997). Copper-gold interaction during the processing of copper-gold ores. *World Gold '97 Conference*, Singapore, September 1 - 3.

Turton, R., Bailie, R.C., Whiting, W.B. and Shaeiwitz, J.A. (2010). *Analysis, Synthesis, and Design of Chemical Processes* (3rd Edition). Prentice Hall, Boston, USA.

van der Merwe, P.F. (1991). Fundamentals of the elution of gold from activated carbon. *PhD Thesis*, Department of Process Engineering, Stellenbosch University, South Africa.

van Deventer, J., Wyethe, J.P., Kotze, M.H. and Shannom, J. (2000). *Journal of the South African Institute of Mining and Metallurgy*, vol. 100, no. 4, pp. 221 - 228.

van Deventer, J.S.J. (1984). Kinetic model for the adsorption of metal cyanides on activated charcoal. *PhD Thesis*, Department of Process Engineering, Stellenbosch University, South Africa.

van Deventer, J.S.J. and van der Merwe, P.F. (1994). Factors affecting the elution of gold cyanide from activated carbon. *Minerals Engineering*, vol. 7, no. 1, pp. 71 - 86.

van Tonder, C. and Sole, M.J. (2006). Innovations in precious metal recovery. *Journal of the South African Institute of Mining and Metallurgy*, vol. 106, no. 4, pp. 297 - 304.

Vermaak, M.K.G. (2005). Fundamentals of the flotation behaviour of palladium bismuth tellurides. *PhD Thesis*, University of Pretoria.

Walton, R. (2005). Zinc Cementation. *In: Adams, M.D. (Ed.), Developments in Mineral Processing,* Vol. 15, Elsevier, Amsterdam, The Netherlands, pp. 589 - 601.

White, J.A. (1994). The Potgietersrus prospect - Geology and exploration history. *Proceedings of the XVth CMMI Congress, Geology, 3, The South African Institute of Mining and Metallurgy*, Johannesburg, South Africa, pp. 173 - 182.

Whyte, R.M., Dempsey, P. and Stange, W. (1990). The development and testing of the AAC Pump-Cell at Vaal Reefs Exploration and Mining Company Limited. *International Reef Mining Conference: Innovations in Metallurgical Plants* Johannesburg, South Africa.

Williams, D.F. and Glasser, D. (1985). The modelling and simulation of processes for the adsorption of gold by activated charcoal. *Journal of the South African Institute of Mining and Metallurgy*, vol. 85, no. 8, pp. 237 - 243.

Woollacott, L.C., Stange, W. and King, R.P. (1990). Towards more effective simulation of CIP and CIL processes. 1. The modelling of adsorption and leaching. *Journal of the South African Institute of Mining and Metallurgy*, vol. 90, no. 10, pp. 275 - 282.

Zanitsch, R.H. and Lynch, R.T. (1978). Selecting a thermal regeneration system for activated carbon. *Chemical Engineering*, vol. 85, no 1, pp. 95 - 100.

Appendices

Appendix A: Process Flow Diagrams

The process flow diagrams are presented as follows:

- As all flowsheet alternatives start with a SART plant, only one SART plant process flow diagram is presented. This general flow diagram are the same for all the different flowsheet alternatives. The only difference is the feed to each process. (Figure A.1)
- All adsorption sections, both activated carbon and ion exchange resins, are generally the same and only differ as a result of the number of adsorption tanks required and adsorbent used. As a result of this, a general adsorption process flow diagram is presented. Although the carousel mode of operation was adopted this diagram illustrates the counter-current nature of the adsorption process. (Figure A.2)
- The elution, recovery and regeneration circuits of the different flowsheet alternatives are, however, different due to the multiple eluting agents, multiple resins and different regeneration cycles between activated carbon and ion exchange resins. Elution, recovery and regeneration process flow diagrams are thus presented for each of the different scenarios, as listed below:
    - Carbon elution and regeneration (CIS 1 and CIS 2 primary circuits) (Figure A.3)
    - Carbon elution and regeneration (CIS 2 secondary circuit) (Figure A.4)
    - Resin elution and regeneration (RIS 1 and RIS 2 primary circuits) (Figure A.5)
    - Resin elution and regeneration (RIS 2, RIS 3, RIS 4 and RIS 5 secondary circuits) (Figure A.6)
    - Resin elution and regeneration (RIS 3 and RIS 4 primary circuits) (Figure A.7)
    - Resin elution (RIS 5 primary circuit) (Figure A.8)

  These different elution, recovery and regeneration process flow diagrams are the same for the different flowsheet alternatives listed in brackets.
- The full process flow diagram of each of these different flowsheet alternatives would thus be the combined SART, adsorption and elution sections, as required for each process.

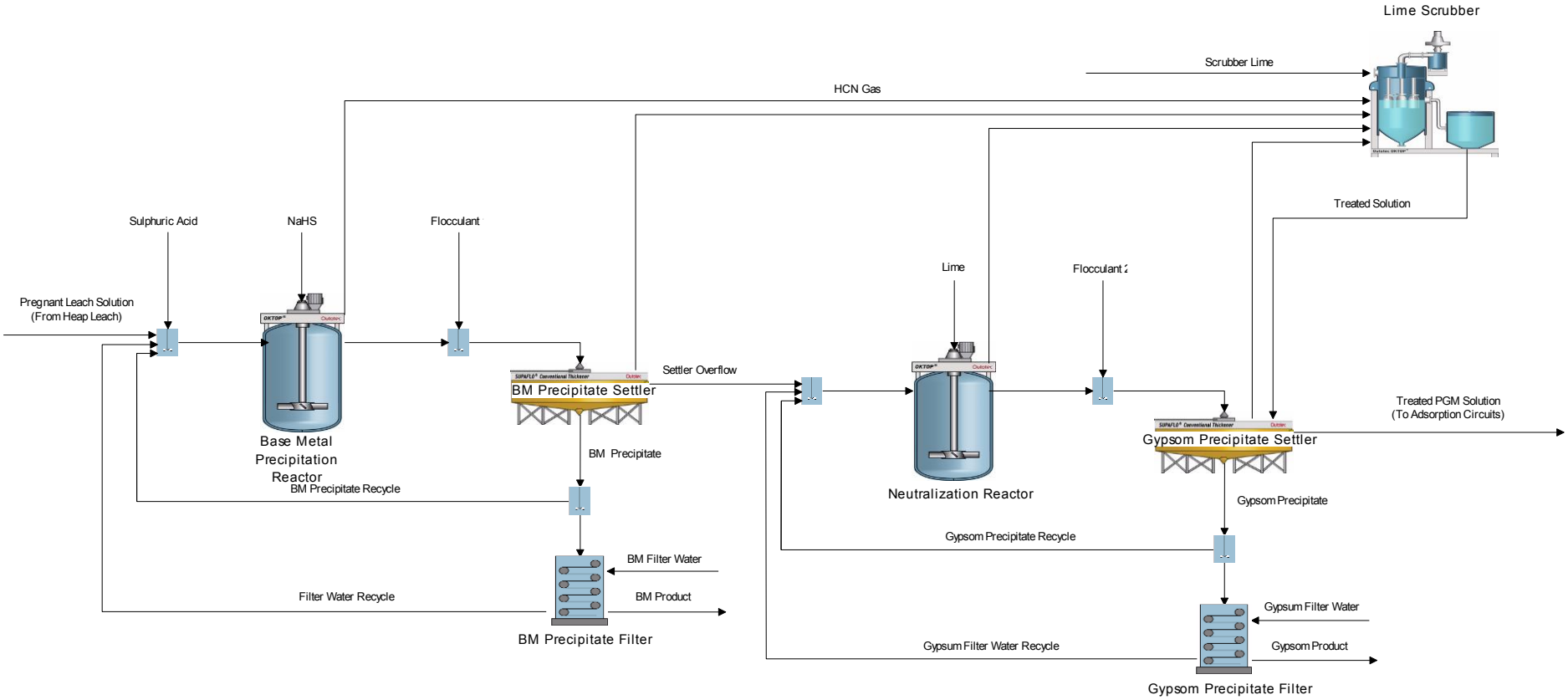

Figure A.1: SART Process Flow Diagram (Primary and Secondary Circuits)

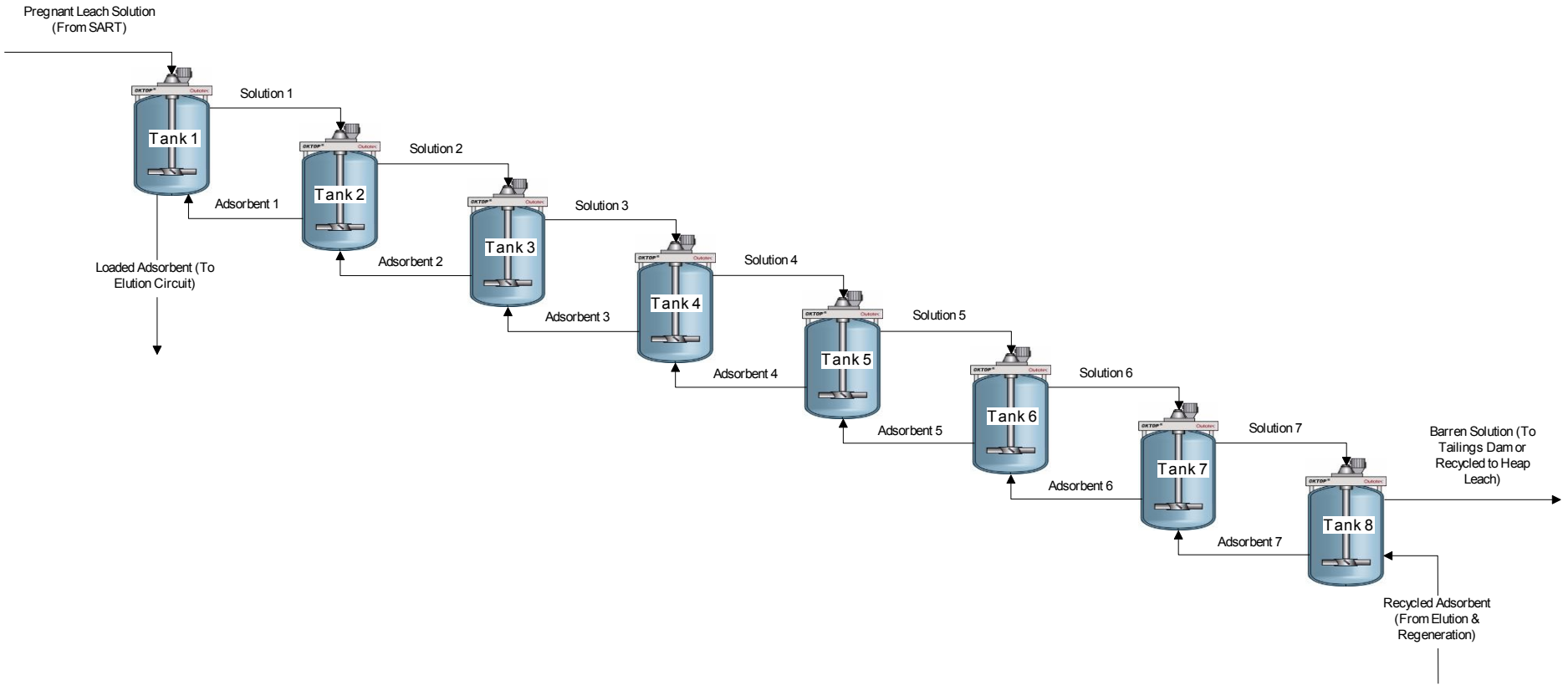

Figure A.2: Adsorption Circuit General Process Flow Diagram (RIS and CIS, Primary and Secondary Circuits)

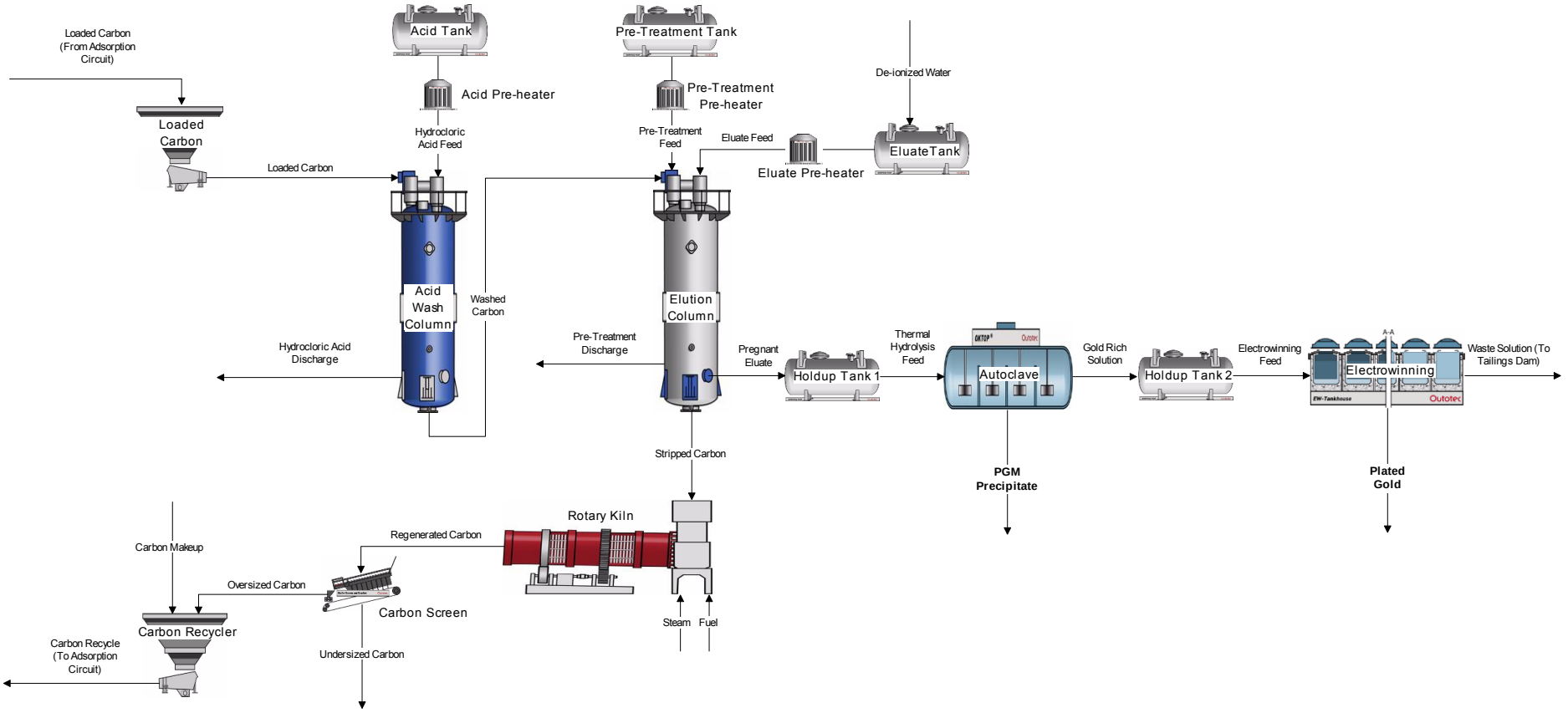

Figure A.3: *Carbon Elution and Regeneration Process Flow Diagram (CIS 1 and CIS 2 Primary Circuits)*

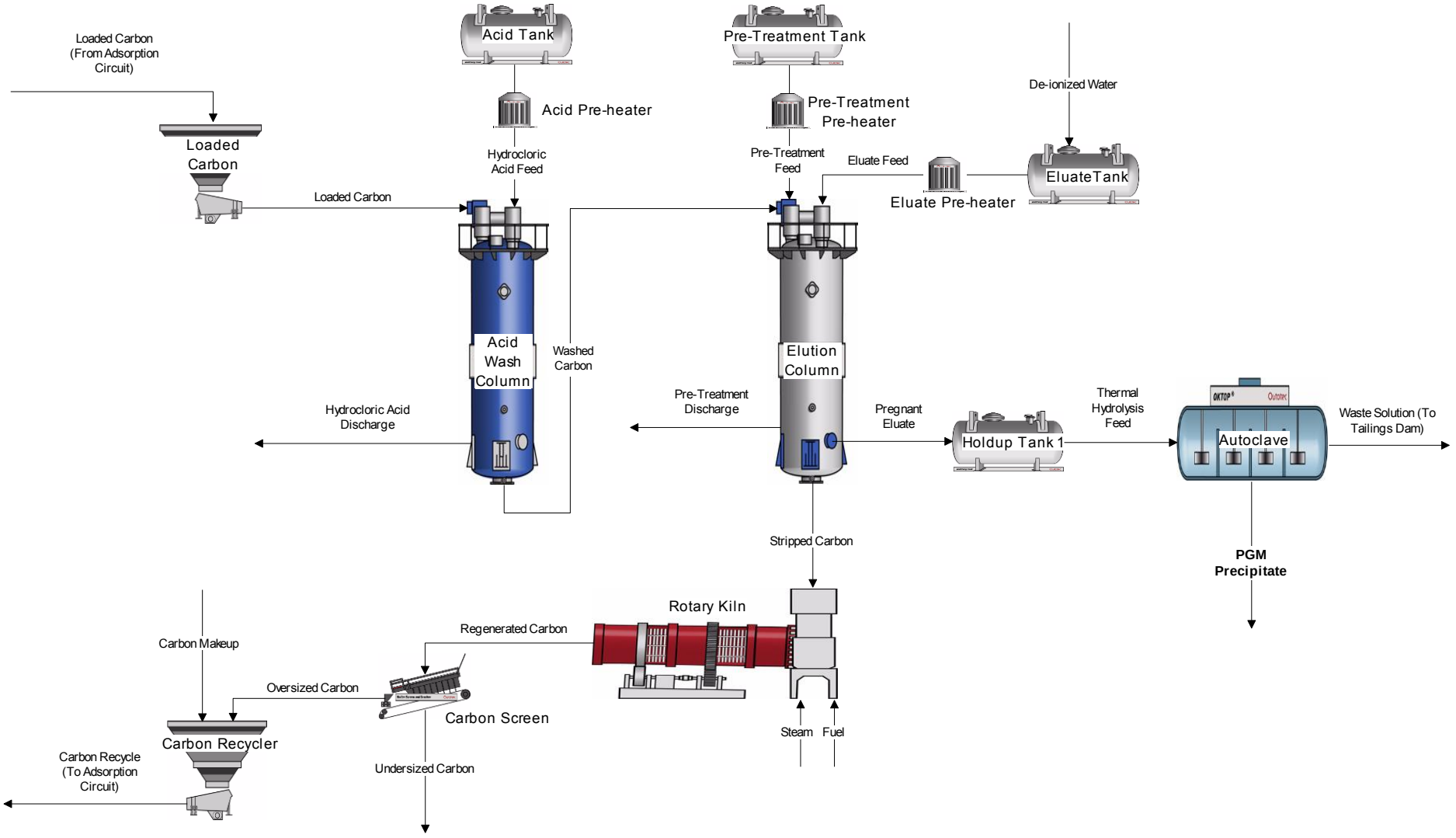

Figure A.4: Carbon Elution and Regeneration Process Flow Diagram (CIS 2 Secondary Circuit)

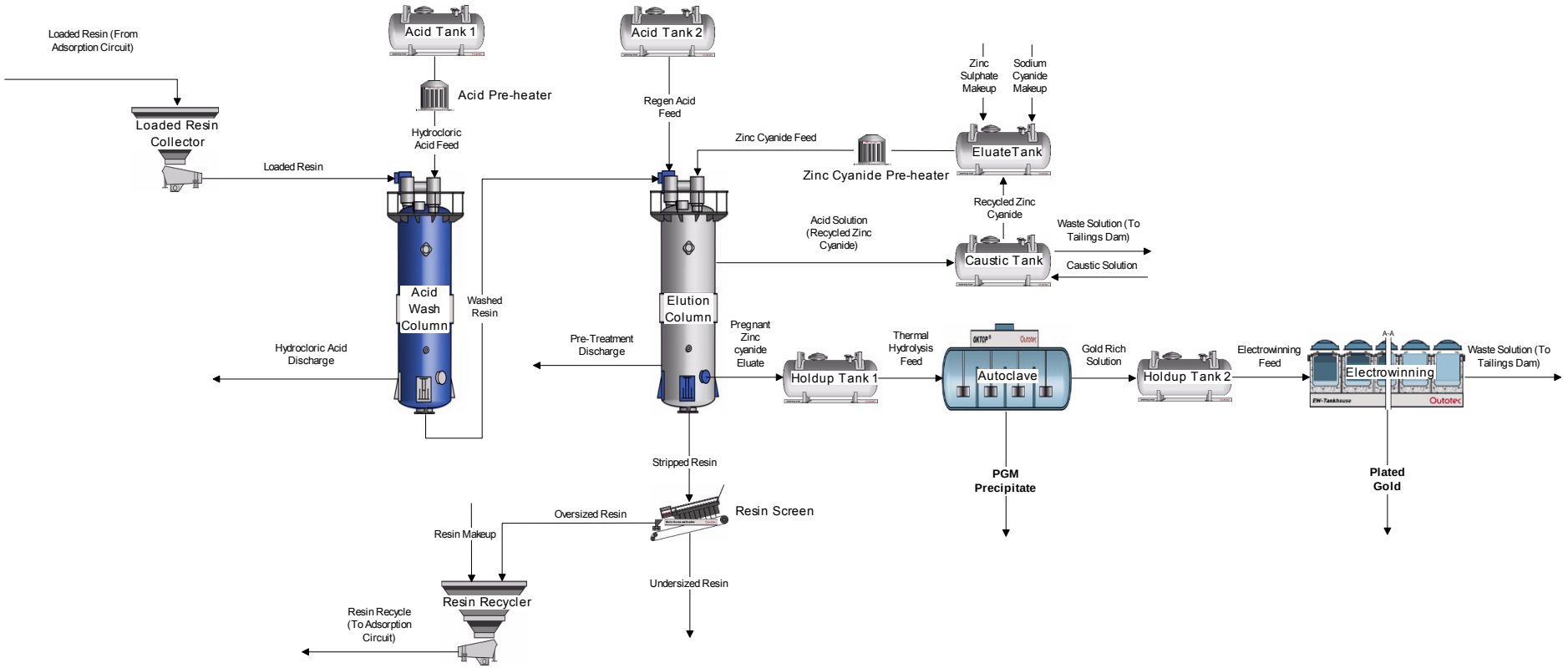

Figure A.5: Resin Elution and Regeneration Process Flow Diagram (RIS 1 and RIS 2 Primary Circuits)

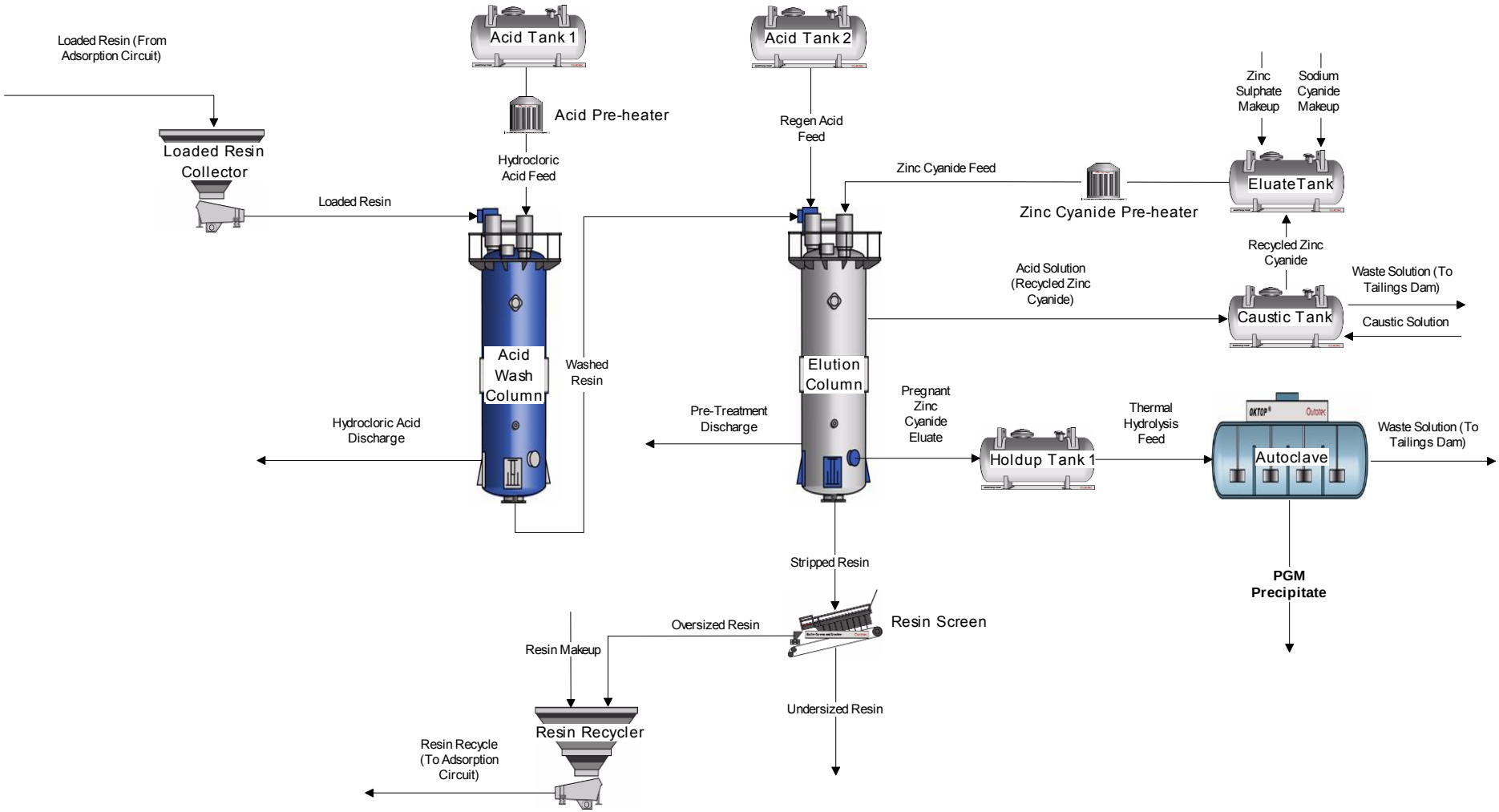

Figure A.6: Resin Elution and Regeneration Process Flow Diagram (RIS 2, RIS 3, RIS 4 and RIS 5 Secondary Circuits)

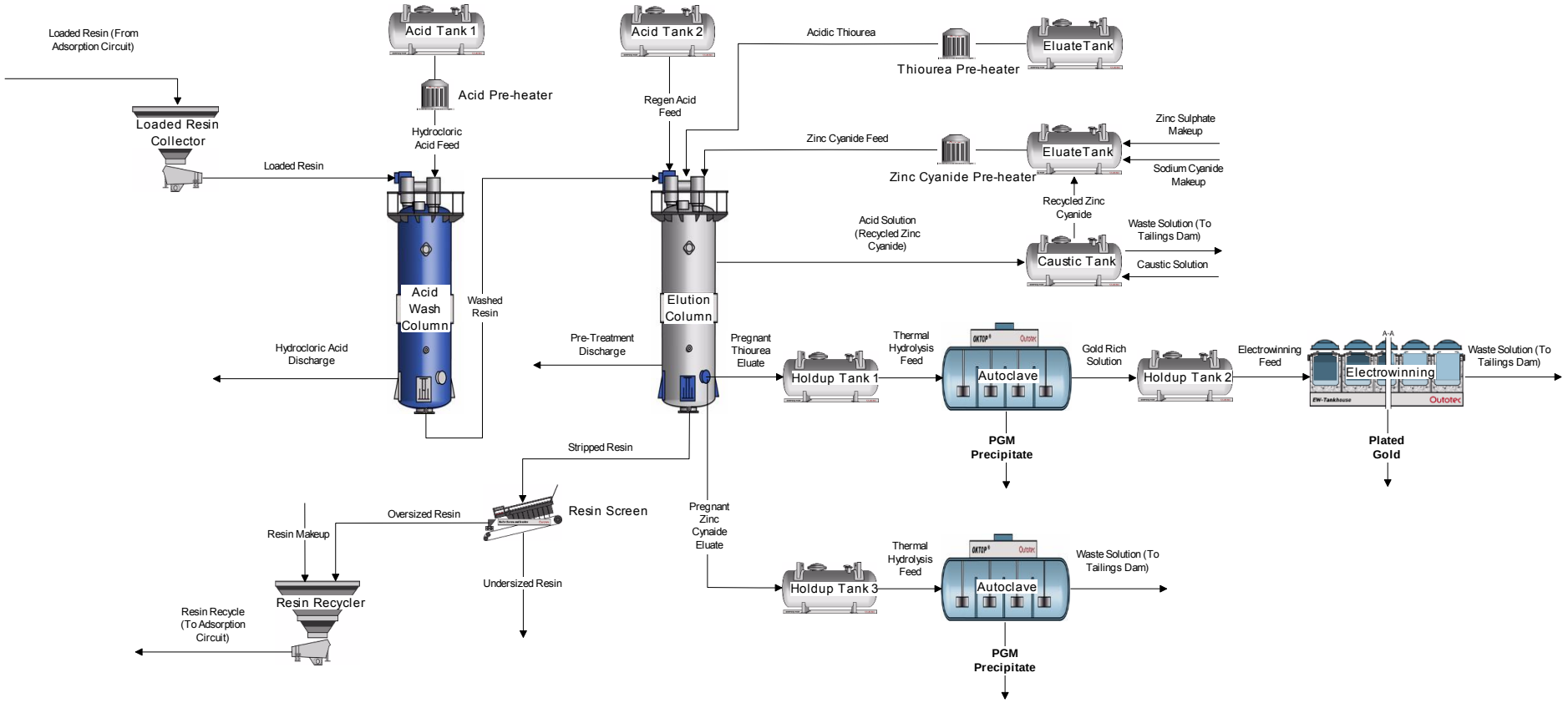

Figure A.7: Resin Elution and Regeneration Process Flow Diagram (RIS 3 and RIS 4 Primary Circuits)

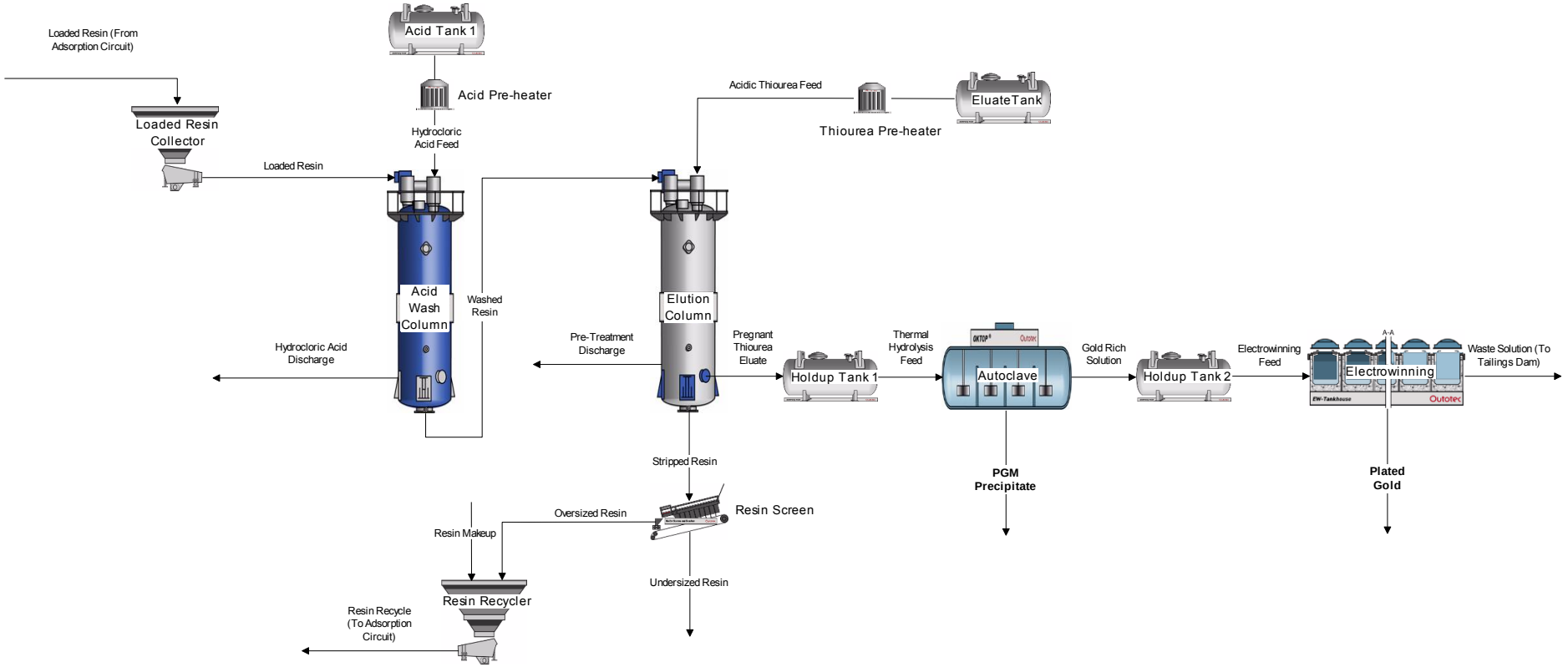

Figure A.8: *Resin Elution Process Flow Diagram (RIS 5 Primary Circuit)*

Appendix B: Equipment Design and Sizing

Equipment design and sizing are presented as follows:

- Detailed sizes, materials of construction and operating conditions, temperature and pressure, for each of the different process units:
    - Carbon adsorption circuits (Table B.1)
    - Resin adsorption circuits (Table B.2)

Table B.1: Equipment sizes and operating conditions (CIS flowsheets)

Process Unit	Unit Size				Design Conditions		
	CIS 1	CIS 2a	CIS 2b	Unit	Material of Construction	Temperature (°C)	Pressure (Bar)
SART							
Plant Section	8498	1418	7080	m^3/d		†	
Adsorption							
Pump Cells	122.37	20.57	101.69	m^3		†	
Collection Tank	11.48	3.58	7.57	m^3	CS	25	1
Elution and Recovery							
Acid Wash Column	10.57	3.30	6.97	m^3	H	90	1
Elution Column	10.57	3.30	6.97	m^3	H	120	3
Acid Tank	9.57	2.99	6.31	m^3	H	25	1
Pre-treatment Tank	9.57	2.99	6.31	m^3	SS	25	1
Eluate Tank	97.38	60.03	50.24	m^3	SS	25	1
Hold-up tank (no 1)	97.38	60.03	50.24	m^3	SS	120	1
Autoclave	6.15	3.79	2.51	m^3	CS	250	40
Hold-up tank (no 2)	7.38	4.55	3.01	m^3	SS	250	1
Electrowinning Cells	5.12	3.15	2.09	t/h	SS	80	1
Regeneration							
Rotary Kiln	5.02	1.56	3.32	t/d		†	
Conditioning Tank	9.66	3.00	6.38	m^3	CS	25	1

a – Primary Plant Section; b – Secondary Plant Section

† Whole plant section – Costs independent of operating conditions

CS – Carbon Steel; SS – Stainless Steel 316; H – Hastelloy C

Table B.2: Equipment sizes and operating conditions (RIS flowsheets)

Process Unit	*Unit Size*									*Design Conditions*		
	RIS 1	*RIS 2a*	*RIS 2b*	*RIS 3*	*RIS 4a*	*RIS 4b*	*RIS 5a*	*RIS 5b*	*Unit*	*MoC*	*Temperature (°C)*	*Pressure (bar)*
SART												
Section	8498	1418	7080	8498	1418	7080	1418	7080	m^3/d		†	
Adsorption												
Pump Cells	60.46	10.17	50.14	60.42	10.22	50.14	10.18	60.44	m^3		†	
Collection Tank	3.64	1.60	2.46	2.64	0.91	2.46	0.45	2.75	m^3	CS	25	1
Elution and Recovery												
Acid Wash Column	3.45	1.52	2.33	2.50	0.86	2.33	0.42	2.61	m^3	H	90	1
Elution Column	3.45	1.52	2.33	2.50	0.86	2.33	0.42	2.61	m^3	H	60	1
Acid Tank	24.23	10.68	16.40	17.58	6.08	16.40	2.97	18.32	m^3	H	25	1
Eluate Tank	45.44	20.01	30.75	32.96	11.39	30.75	5.57	34.36	m^3	SS	25	1
Hold-up tank (no 1)	45.44	20.01	30.75	32.96	11.39	30.75	5.57	34.36	m^3	SS	60	1
Autoclave (no 1)	2.87	1.26	1.54	2.08	0.72	1.54	0.35	1.72	m^3	CS	250	1
Hold-up tank (no 2)	3.44	1.52	1.84	3.75	1.30	1.84	0.42	2.06	m^3	SS	250	40
Electrowinning Cells	2.26	1.00	1.21	1.64	0.57	1.21	0.28	1.35	t/h	SS	80	1
Hold-up tank (no 3)				32.96	11.39				m^3	SS	60	1
Autoclave (no 2)				1.65	0.57				m^3	CS	250	40
Regeneration												
Acid Tank	5.17	2.28	3.50	3.75	1.30	3.50		3.91	m^3	H	25	1
Caustic Soda Tank	11.25	4.95	7.61	8.16	2.82	7.61		8.51	m^3	SS	25	1
Conditioning Tank	3.03	1.33	2.05	2.20	0.76	2.05	0.37	2.29	m^3	CS	25	1

a – Primary plant section; b – Secondary Plant Section

† Whole plant section – Costs independent of operating conditions

MoC – Material of Construction; CS – Carbon Steel; SS – Stainless Steel 316; H – Hastelloy C

Appendix C: Capital Cost Estimates

Capital cost estimates are presented as follows:

- Equipment base case costs and correction factors, as obtained from literature (Table C.1 and Table C.2)
- Estimated equipment capital costs for each flowsheet alternative (Table C.3 and Table C.4)
- Capital cost breakdown for each flowsheet alternative (Table C.5)
- Capital cost breakdown per plant section for each flowsheet alternative (Table C.6)
- Miscellaneous capital costs for each flowsheet alternative (Table C.7). These includes:
    - Fixed capital investment
    - Working capital
    - Inventory costs
    - Value material lock-up

Table C.1: Equipment base case costs and correction factors (CIS flowsheets)

Process Unit	Base Case Costs					Correction Factors			Scaling Factor (n)
	Size	Unit	Cost (US$)	Year	Reference	F_M	F_T	F_P	
SART									
Section†	1	m^3/d	1 200	2012	Stewart & Kappes (2012)	1	1	1	0.60
Adsorption									
Pump Cells‡	30	m^3	216 837	2013	Kemix (2013)	1	1	1	0.47
Collection Tank	5	m^3	11 500	2000	Gerrard (2000)	1	1	1	0.53
Elution and Recovery									
Acid Wash Column	2.6	m^3	5 366	2001	Turton *et al.* (2010)	5.8	1	1	0.30
Elution Column	2.6	m^3	5 366	2001	Turton *et al*. (2010)	5.8	1	1	0.30
Acid Pre-heater	1500	kW	52 451	2001	Turton *et al*. (2010)	1	1	1	0.81
Pre-Treatment Pre-heater	1500	kW	52 451	2001	Turton *et al*. (2010)	1	1	1	0.81
Eluate Pre-heater	1500	kW	52 451	2001	Turton *et al*. (2010)	1	1	1	0.81
Acid Tank	5	m^3	11 500	2000	Gerrard (2000)	5.8	1	1	0.53
Pre-Treatment Tank	5	m^3	11 500	2000	Gerrard (2000)	1.8	1	1	0.53
Eluate Tank	5	m^3	11 500	2000	Gerrard (2000)	1.8	1	1	0.53
Hold-up tank (no 1)	5	m^3	11 500	2000	Gerrard (2000)	1.8	1.6	1	0.53
Autoclave	6	t	98 400	2000	Gerrard (2000)	1	1.6	1.9	0.82
Hold-up tank (no 2)	5	m^3	11 500	2000	Gerrard (2000)	1.8	1.6	1	0.53
Electrowinning Cells	5	t/h	42 092	2013	Kemix (2013)	1	1	1	0.80
Regeneration									
Rotary Kiln	4.55	t/d	360 000	1978	Zanitsch & Lynch (1978)	1	1	1	0.70
Conditioning Tank	5	m^3	11 500	2000	Gerrard (2000)	1.8	1	1	0.53

† Total plant cost

‡ Cost include that of design and engineering, piping, installation and commissioning

Table C.2: Estimated equipment capital costs (CIS flowsheets)

Process Unit	*Unit Costs (US$)*		
	CIS 1	*CIS 2a*	*CIS 2b*
SART			
Section†	10 370 234	3 541 865	9 294 220
Adsorption			
Pump Cells‡	3 404 949	1 465 219	3 115 747
Collection Tank	27 205	14 676	21 819
Elution and Recovery			
Acid Wash Column	96 329	52 747	76 728
Elution Column	154 127	84 395	122 765
Acid Wash Pre-heater	54 228	21 988	39 431
Pre-Treatment Pre-Heater	7 430	3 012	5 401
Eluate Pre-heater	121 680	50 602	89 217
Acid Tank	152 038	77 278	114 896
Pre-treatment Tank	44 459	23 983	35 657
Eluate Tank	152 038	117 656	107 056
Hold-up tank (no 1)	243 261	188 249	171 290
Autoclave	463 998	312 073	222 651
Hold-up tank (no 2)	61 982	47 965	
Electrowinning Cells	42 670	29 110	
Regeneration			
Rotary Kiln	1 043 614	460 579	780 748
Conditioning Tank	24 824	13 363	19 927

† Total plant cost

‡ Cost include that of design and engineering, piping, installation and commissioning

Table C.3 *Equipment base case costs and correction factors (RIS flowsheets)*

Process Unit	Base Case Costs					Correction Factors			Scaling Factor (n)
	Size	Unit	Cost (US$)	Year	Reference	F_M	F_T	F_P	
SART									
Section†	1	m^3/d	1 200	2012	Stewart & Kappes (2012)	1	1	1	0.60
Adsorption									
Pump Cells‡	30	m^3	216 837	2013	Kemix (2013)	1	1	1	0.47
Collection Tank	5	m^3	11 500	2000	Gerrard (2000)	1	1	1	0.53
Elution and Recovery									
Acid Wash Column	2.6	m^3	5 366	2001	Turton *et al*. (2010)	5.8	1	1	0.30
Elution Column	2.6	m^3	5 366	2001	Turton *et al*. (2010)	5.8	1	1	0.30
Acid Pre-heater	1500	kW	52 451	2001	Turton *et a*l. (2010)	1	1	1	0.81
Eluate Pre-heater	1500	kW	52 451	2001	Turton *et al*. (2010)	1	1	1	0.81
Acid Tank	5	m^3	11 500	2000	Gerrard (2000)	5.8	1	1	0.53
Eluate Tank	5	m^3	11 500	2000	Gerrard (2000)	1.8	1	1	0.53
Hold-up tank (no 1)	5	m^3	11 500	2000	Gerrard (2000)	1.8	1	1	0.53
Autoclave (no 1)	6	t	98 400	2000	Gerrard (2000)	1	1.6	1.9	0.82
Hold-up tank (no 2)	5	m^3	11 500	2000	Gerrard (2000)	1.8	1.6	1	0.53
Electrowinning Cells	5	t/h	42 092	2013	Kemix (2013)	1	1	1	0.80
Hold-up tank (no 3)	5	m^3	11 500	2000	Gerrard (2000)	1.8	1.6	1	0.53
Autoclave (no 2)	6	t	98 400	2000	Gerrard (2000)	1	1.6	1.9	0.82
Regeneration									
Acid Tank	5	m^3	11 500	2000	Gerrard (2000)	5.8	1	1	0.53
Caustic Soda Tank	5	m^3	11 500	2000	Gerrard (2000)	1.8	1	1	0.53
Conditioning Tank	5	m^3	11 500	2000	Gerrard (2000)	1.8	1	1	0.53

† *Total plant cost*

‡ *Cost include that of design and engineering, piping, installation and commissioning*

Table C.4: Estimated equipment capital costs (RIS flowsheets)

Process Unit	Unit Costs (US$)							
	RIS 1	RIS 2a	RIS 2b	RIS 3	RIS 4a	RIS 4b	RIS 5a	RIS 5b
SART								
Section†	10 370 234	3 541 865	9 294 220	10 370 235	3 541 865	9 294 220	3 541 865	9 294 220
Adsorption								
Pump Cells‡	2 418 109	909 164	1 939 197	2 417 334	911 075	1 939 197	1 039 058	2 115 591
Collection Tank	14 786	9 573	12 022	12 473	7 103	12 022	4 862	12 750
Elution and Recovery								
Acid Wash Column	53 869	37 467	45 019	46 441	30 110	45 019	23 643	47 319
Elution Column	53 869	37 467	45 019	46 441	30 110	45 019	23 643	47 319
Acid Wash Pre-heater	64 967	34 616	48 227	50 868	22 287	48 227	12 617	52 508
Eluate Pre-heater (no 1)	35 554	18 688	26 222	35 983	15 607	26 222	8 776	28 603
Eluate Pre-heater (no 2)				27 690	11 920			
Acid Tank (no 1)	237 399	151 752	190 573	197 731	112 601	190 573	77 075	202 126
Eluate Tank (no 1)	101 506	65 716	82 527	85 627	48 761	82 527	33 377	87 530
Eluate Tank (no 2)				85 627	48 761			
Hold-up tank (no 1)	101 506	65 716	82 527	85 627	48 761	82 527	33 377	87 530
Autoclave (no 1)	237 679	121 298	142 466	182 677	76 445	142 466	42 525	156 049
Hold-up tank (no 2)	25 863	16 744		21 817	12 424		8 504	
Electrowinning Cells	22 318	11 578		17 264	7 380		4 164	
Hold-up tank (no 3)				85 627	48 761			
Autoclave (no 2)				150 831	63 118			
Regeneration								
Acid Tank (no 2)	103 390	66 935	84 059	87 216	49 666	84 059		89 154
Caustic Soda Tank	48 438	31 359	39 381	40 860	23 269	39 381		41 469
Conditioning Tank	13 431	8 693	10 918	11 331	6 452	10 918	4 418	11 582

† Total plant cost

‡ Cost include that of design and engineering, piping, installation and commissioning

Table C.5: Capital cost breakdown for each flowsheet alternative (CIS and RIS)

Expense	*Capital Cost (US$)*						
	CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
Direct Costs							
Purchased Equipment Delivered	7 866 138	9 979 732	5 640 829	6 999 874	5 801 782	6 987 933	6 847 953
Equipment Installation	2 978 440	3 778 734	2 135 848	2 650 438	2 196 791	2 645 916	2 592 914
Piping (Installed)	2 367 478	3 003 609	1 697 725	2 106 758	1 746 168	2 103 164	2 061 034
Instrumentation & Controls	992 813	1 259 578	711 949	883 479	732 264	881 972	864 305
Electrical (Installed)	763 703	968 906	547 653	679 599	563 280	678 440	664 850
Utilities	2 672 959	3 391 171	1 916 786	2 378 598	1 971 479	2 374 540	2 326 974
Off-sites	1 527 405	1 937 812	1 095 307	1 359 199	1 126 560	1 356 880	1 329 700
Buildings (including services)	2 214 738	2 809 827	1 588 194	1 970 838	1 633 512	1 967 476	1 928 065
Site Preparation	458 222	581 344	328 592	407 760	337 968	407 064	398 910
Total Direct Cost	21 841 897	27 710 712	15 662 883	19 436 543	16 109 804	19 403 387	19 014 705
Indirect Costs							
Design, Engineering and Supervision	2 443 849	3 100 499	1 752 490	2 174 718	1 802 495	2 171 008	2 127 519
Construction Expenses	2 596 589	3 294 280	1 862 021	2 310 638	1 915 151	2 306 696	2 260 489
Total Indirect Cost	5 040 438	6 394 780	3 614 511	4 485 356	3 717 647	4 477 705	4 388 009
Contractors Fee	1 344 117	1 705 275	963 870	1 196 095	991 373	1 194 055	1 170 136
Contingency	6 720 584	8 526 373	4 819 349	5 980 475	4 956 863	5 970 273	5 850 679
Total fixed capital cost	34 947 034	44 337 140	25 060 613	31 098 470	25 775 686	31 045 420	30 423 528
Working Capital	6 167 124	7 824 201	4 422 461	5 487 965	4 548 650	5 478 603	5 368 858
Total Capital Cost	41 114 158	52 161 341	29 483 074	36 586 435	30 324 336	36 524 023	35 792 386

Table C.6: Capital cost per plant section for each flowsheet alternative (CIS and RIS)

Plant Section	Fixed Capital Investment (US$)						
	CIS 1	CIS 2	RIS 1	RIS 2	RIS 3	RIS 4	RIS 5
SART	13 481 305	16 686 910	13 481 305	16 686 910	13 481 305	16 686 910	16 686 910
Adsorption	9 386 172	12 628 015	6 706 584	7 903 124	6 698 070	7 901 584	8 744 740
Elution	7 332 789	9 359 450	4 138 529	5 436 207	4 976 961	5 507 318	4 339 140
Regeneration	4 746 769	5 662 765	734 196	1 072 228	619 350	949 607	652 739
Total	34 947 034	44 337 140	25 060 613	31 098 470	25 775 686	31 045 420	30 423 528

Table C.7: Miscellaneous capital costs for each flowsheet alternative (CIS and RIS)

Plant Section	Miscellaneous Capital Costs (US$)						
	CIS 1	CIS 2	RIS 1	RIS 2	RIS 3	RIS 4	RIS 5
Fixed Investment	34 947 034	44 337 140	25 060 613	31 098 470	25 775 686	31 045 420	30 423 528
Working Capital	6 167 124	7 824 201	4 422 461	5 487 965	4 548 650	5 478 603	5 368 858
Carbon Inventory	19 976	19 170	-	-	-	-	-
Resin Inventory	-	-	63 806	43 282	60 086	46 803	67 208
Locked-up Material	19 934 570	14 655 417	16 953 776	9 508 329	19 298 843	13 154 676	19 224 998
Total	61 068 704	66 835 928	46 500 656	46 138 046	49 683 265	49 725 502	55 084 592

Appendix D: Operating Cost Estimates

Operating cost estimates are presented as follows:

- Raw material requirements and costs (Table D.1 and Table D.2).
- Waste treatment requirements and costs (Table D.3 and Table D.4).
- Utility requirements and costs (Table D.5 and Table D.6).
- Operating cost breakdown for each flowsheet alternative (Table D.7).
- Operating cost per plant section for each flowsheet alternative (Table D.8).

Table D.1: *Raw material requirements for each flowsheet alternative (CIS and RIS)*

Raw Material	*Purity (%)*	*Flowrate (t/a)*						
		CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
Sulphuric Acid	40	14 168	14 168	14 168	14 168	14 168	14 168	14 168
Sodium Hydrosulphate	70	497	497	497	497	497	497	497
Flocculant (Hychem AF 305)	100	2,68E-01	2,68E-01	2,68E-01	2,68E-01	2,68E-01	2,68E-01	2,68E-01
Lime	20	19 468	19 461	19 468	19 461	19 468	19 461	19 461
Scrubber Lime	5	12 510	12 520	12 510	12 520	12 510	12 520	12 520
Hydrochloric Acid	6	3 219	3 127	8 150	9 104	5 913	7 559	7 162
Hydrochloric Acid	40			1 743	1 947	1 265	1 617	1 318
Hydrochloric Acid (*Thiourea Elution*)	100					72	25	12
De-ionized Water	100	32 752	37 088					
Carbon Makeup	-	117	113					
Resin Makeup	-			7,69	8,31	10,83	8,78	7,74
Wash Water	100	5 199	5 108	2 999	6 258	2 719	2 925	2 876
Process Water	100			10 144	11 331	17 266	12 832	9 345
Caustic Soda	20			3 791	4 235	2 750	3 516	2 867
Sodium Cyanide	100			188	210	136	174	142
Zinc Sulphate	100			155	173	112	144	117
Cyanide Pre-treatment	99	3 219	3 127					
Thiourea	100					529	183	89

Table D.2: Raw material cost per annum for each flowsheet alternative (CIS and RIS)

Raw Material	*Unit Cost ($US/t)*	*Raw Material Cost ($US/a)*						
		CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
Sulphuric Acid	148	2 094 359	2 094 382	2 094 359	2 094 382	2 094 359	2 094 382	2 094 382
Sodium Hydrosulphate	857	426 036	426 063	426 036	426 063	426 036	426 063	426 063
Flocculant (Hychem AF 305)	1 800	483	483	483	483	483	483	483
Lime	29,6	575 582	575 378	575 582	575 378	575 582	575 378	575 378
Scrubber Lime	8,50	106 276	106 362	106 276	106 362	106 276	106 362	106 362
Hydrochloric Acid	23,2	74 684	72 561	189 105	211 224	137 184	175 380	166 184
Hydrochloric Acid	148			257 673	287 813	186 926	238 971	194 843
Hydrochloric Acid (Thiourea Elution)	364					26 323	9 098	4 450
De-ionized Water	4,42	144 819	163 992					
Carbon Makeup	1 920	223 906	217 174					
Resin Makeup	18 829			144 881	156 452	203 965	165 243	145 724
Wash Water	1,47	7 655	7 520	4 416	9 214	4 004	4 307	4 234
Process Water	1,47			14 936	16 683	25 421	18 893	13 759
Caustic Soda	135			509 992	569 645	369 969	472 977	385 637
Sodium Cyanide	1 347			253 136	282 745	183 635	234 763	191 412
Zinc Sulphate	866			134 028	149 705	97 229	124 300	101 347
Cyanide Pre-treatment	32,0	103 132	100 200					
Thiourea	2 000					1 057 895	365 644	178 832
Total		3 756 931	3 764 115	4 710 902	4 886 149	5 495 287	5 012 245	4 589 089

Table D.3: Waste stream classification and flowrate for each flowsheet alternative (CIS and RIS)

Waste Stream	*Classification*	*Flowrate (t/a)*						
		CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
Hydrochloric Acid Discharge	2	6 445	6 262	16 317	18 224	11 840	15 133	14 341
Pre-Treatment Discharge	2	3 219	3 127					
Waste Solution (Barren Eluate)	1	32 764	37 100	13 622	15 214	20 393	16 266	12 079
Zinc Cyanide Waste	2			1 045	1 167	758	969	790
Total		42 428	46 489	30 983	34 605	32 991	32 368	27 209

Table D.4: Waste treatment cost for each flowsheet alternative (CIS and RIS)

Waste Stream	*Treatment Cost (US$/t)*	*Treatment Cost (US$/a)*						
		CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
Hydrochloric Acid Discharge	104	669 803	650 779	1 695 669	1 893 846	1 230 465	1 572 684	1 490 277
Pre-Treatment Discharge	104	334 504	324 994					
Waste Solution (Barren Eluate)	41	1 346 958	1 525 222	560 005	625 463	838 379	668 711	496 584
Zinc Cyanide Waste	104			108 571	121 270	78 761	100 691	82 097
Total		2 351 265	2 500 995	2 364 244	2 640 579	2 147 605	2 342 085	2 068 958

Table D.5: Utility requirements for each flowsheet alternative (CIS and RIS)

Utility	*Unit*	*Utility Requirements*						
		CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
Boiler Steam								
Acid Pre-heaters	GJ/a	1 095	1 064	2 772	3 096	2 011	2 571	2 436
Pre-Treatment Pre-heaters	GJ/a	1 600	1 555					
Elute Pre-heaters	GJ/a	16 280	18 436	1 895	2 117	3 299	2 423	1 758
Rotary Kiln								
Fuel	GJ/a	44 890	43 609					
High Pressure Steam	t/a	1 693	1 645					
Power								
Autoclave	kWh/a	8 092 644	9 164 041	4 210 880	3 997 909	5 320 342	4 260 908	3 674 677
Electrowinning	kWh/a	57 400	56 367	45 384	50 874	49 504	53 968	10 576
Electricity								
Plant Requirement	MWh/a	49 521	57 935	42 679	49 860	45 491	49 323	46 302

- Boiler steam and autoclave power were determined with the use of energy balances.
- Fuel and steam requirements for the rotary kiln was obtained from Zanitsch & Lynch (1978).
    - Fuel requirements, 18.61 GJ/t carbon
    - Steam requirements, 1 t/t carbon
- Plant electricity requirements were estimated as 20% of the total operating costs (Peters & Timmerhaus, 1991).

Table D.6: Utility costs for each flowsheet alternative (CIS and RIS)

Utility	*Cost*	*Utility Requirements Cost (US$/a)*						
		CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
Boiler Steam								
Acid Pre-heaters	20,35 $/GJ	22 277	21 644	56 407	63 005	40 920	52 313	49 570
Pre-Treatment Pre-heaters	20,35 $/GJ	32 559	31 633					
Elute Pre-heaters	20,35 $/GJ	331 287	375 147	38 559	43 069	67 133	49 295	35 777
Rotary Kiln								
Fuel	16,32 $/GJ	732 826	711 910					
High Pressure Steam	34,45 $/t	58 337	56 672					
Power								
Autoclave	0,092 $/kWh	743 202	841 596	386 713	367 155	488 603	391 308	337 470
Electrowinning	0,092 $/kWh	5 271	5 177	4 168	4 672	4 546	4 956	971
Electricity								
Plant Requirement	0,092 $/kWh	4 547 848	5 320 604	3 919 527	4 578 946	4 177 725	4 529 655	4 252 263
Total		6 473 608	7 364 382	4 405 375	5 056 847	4 778 927	5 027 528	4 676 051

- All utility costs obtained from Turton *et al*. (2010).

Table D.7: Operating cost breakdown for each flowsheet alternative (CIS and RIS)

Expense	*Operating Cost (US$)*						
	CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
Direct Operating Costs							
Raw Materials	3 756 931	3 764 115	4 710 902	4 861 198	5 495 287	5 012 045	4 589 089
Waste Treatment	2 351 265	2 500 995	2 364 244	2 640 579	2 147 605	2 342 085	2 068 958
Utilities	6 473 608	7 364 382	4 405 375	5 056 847	4 778 927	5 027 528	4 676 051
Operating Labour	331 700	652 700	331 700	652 700	331 700	652 700	652 700
Direct Supervisory and Labour	59 706	117 486	59 706	117 486	59 706	117 486	117 486
Maintenance and Repairs	2 096 822	2 660 228	1 503 637	1 865 908	1 546 541	1 862 725	1 825 412
Operating Supplies	314 523	399 034	225 546	279 886	231 981	279 409	273 812
Laboratory Charges	49 755	97 905	49 755	97 905	49 755	97 905	97 905
Patents and Royalties	682 177	798 091	587 929	686 842	626 659	679 448	637 839
Total Direct Operating Costs	16 116 487	18 354 936	14 238 793	16 259 351	15 268 161	16 056 416	14 939 252
Fixed Operating Costs							
Local Taxes and Insurance	1 118 305	1 418 788	801 940	995 151	824 822	993 453	973 553
Plant Overhead Costs	1 492 937	2 058 249	1 137 026	1 581 657	1 162 768	1 579 747	1 557 359
Total Fixed Operating Costs	2 611 242	3 477 037	1 938 965	2 576 808	1 987 590	2 573 200	2 530 912
General Operating Expenses							
Administration Costs	373 234	514 562	284 256	395 414	290 692	394 937	389 340
Distribution and Selling Costs	2 501 317	2 926 332	2 155 740	2 518 420	2 297 749	2 491 310	2 338 744
Research and Development	1 136 962	1 330 151	979 882	1 144 736	1 044 431	1 132 414	1 063 066
Total General Operating Costs	4 011 513	4 771 045	3 419 878	4 058 571	3 632 872	4 018 661	3 791 150
Gross Operating Cost	22 739 242	26 603 018	19 597 636	22 894 730	20 888 623	22 648 277	21 261 313
Contingency	5 684 811	6 650 755	4 899 409	5 723 682	5 222 156	5 662 069	5 315 328
Total Operating Cost	28 424 053	33 253 773	24 497 045	28 618 412	26 110 779	28 310 346	26 576 642

Table D.8: *Operating cost per plant section for each flowsheet alternative (CIS and RIS)*

Plant Section	*Operating Costs (US$)*						
	CIS 1	*CIS 2*	*RIS 1*	*RIS 2*	*RIS 3*	*RIS 4*	*RIS 5*
SART	11 143 801	12 685 517	11 143 801	12 685 517	11 143 801	12 685 517	12 685 517
Adsorption	3 540 720	4 788 220	2 280 217	2 929 598	2 195 548	2 847 015	3 181 393
Elution	10 561 286	12 228 316	8 846 529	10 322 255	10 931 185	10 454 679	8 743 756
Regeneration	3 178 246	3 551 720	2 226 498	2 681 043	1 840 244	2 323 135	1 965 977
Total	28 424 053	33 253 773	24 497 045	28 618 412	26 110 779	28 310 346	26 576 641

Appendix E: Alternative Adsorption Models

Although the first order Nicol-Fleming adsorption model was used in the simulation and optimization of the different flowsheets, alternative adsorption models were also investigated based on the findings reported by Mpinga (2012). Mpinga (2012) suggested that a second order model be used in modelling the adsorption of platinum and palladium onto activated carbon, as it was found that a second order model, such as that reported by Ho and McKay (1999), fitted batch adsorption data almost perfectly. This model and a derived second order model based on the first order Nicol-Fleming model are discussed in the next sections, giving insights as to why these models were not used in the simulations.

E.1 Pseudo-second order kinetic adsorption model (Ho & McKay, 1999)

The pseudo-second order model, reported by Ho & McKay (1999), was developed based on the assumption that the adsorption of a specie is directly proportional to the number of active sites available on the adsorbent. Taking this into consideration, a second order rate equation was assumed to hold true. This rate equation is given by equation $(E.1)$:

$$\frac{dq_t}{dt} = k_c(q_e - q_t)^2 \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (E.1)$$

where

q_t = The amount of solute adsorbed on the surface of the adsorbent at t (mg/g)

q_e = The amount of solute adsorbed at equilibrium (mg/g)

k_c = The rate constant of adsorption (g/mg min)

t = Time (min)

Separating the variables and integrating equation $(E.1)$ for the boundary conditions $t = 0$ to $t = t$ and $q_t = 0$ to $q_t = q_t$ gives

$$\frac{1}{(q_e - q_t)} - \frac{1}{q_e} = k_c t \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (E.2)$$

Solving for t yields:

$$t = \frac{q_t}{k_c q_e (q_e - q_t)} \quad \ldots\ldots\ldots\

In deriving the second order kinetic adsorption model, the same general assumptions as reported by Nicol et al. (1984*a*, 1984*b*) in the development of their first order gold adsorption, please refer to section 7.1.5, model was made. The only difference was that it was assumed that the concentration gradients of platinum and palladium within each phase were of a second order nature, as opposed to the linear concentration gradients assumed for gold, this assumption was believe that the valences of the adsorbed ion dictated the order of the adsorption reaction, thus second order for the divalent platinum and palladium cyanide complexes. By applying this assumption, a second order rate equation for the variation of the platinum or palladium concentration on the carbon can be written as:

$$\frac{d[Pt]_c}{dt} = k(K[Pt]_s - [Pt]_c)^2 \quad \ldots\ldots\ldots\

Integrating equation $(E.7)$ for the boundary conditions $t = 0$ to $t = t$ and $[Pt]_c = 0$ to $[Pt]_c = [Pt]_c$ gives:

$$\frac{[Pt]_c}{(-KM_s[Pt]_{s,o})([Pt]_c M_s + [Pt]_c KM_c - KM_s[Pt]_{s,o})} = \frac{k}{{M_s}^2} t \quad \ldots\ldots\ldots\ldots\ldots\ldots \quad (E.8)$$

Simplification of equation $(E.8)$ yields

$$\frac{[Pt]_c}{\beta\,(\beta - [Pt]_c)} = \alpha t \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (E.9)$$

with

$$\alpha = \frac{k(KM_c + M_s)^2}{{M_s}^2} \;; \qquad \beta = \frac{KM_s[Pt]_{s,o}}{(KM_c + M_S)}$$

Solving for $[Pt]_c$:

$$[Pt]_c = \frac{\alpha\beta^2 t}{1 + \alpha\beta t} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (E.10)$$

Rearrangement yields:

$$t = \frac{[Pt]_c}{\alpha\beta(\beta - [Pt]_c)} \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots \quad (E.11)$$

This derived second order model, in the form of equation $(E.10)$, is in a similar form as the pseudo-second order model reported by Ho & McKay (1999) (equation $(E.3)$). This model can be used to determine the rate and equilibrium constants for the adsorption of platinum and palladium onto activated carbon from batch adsorption test data, which can then be used in combining the model with a mass balance over a counter-current adsorption circuit, simulating the circuit performance. Equation $(E.10)$ was found to fit the batch adsorption test data obtained in this study very well, with a coefficient of determination of $R^2 > 0.998$.

The model was then combined with a counter-current adsorption circuit mass balance in order to simulate the different adsorption circuits for each flowsheet alternative. However, the complexity of the model, primarily the quadratic terms, proved to be problematic in integrating the model with the simulator used, as the model requires an iterative approach, similar to that discussed in section 7.4.3, while there are further interdependence between these quadratic terms.

Due to these difficulties in applying the aforementioned models to simulate the performance of an adsorption circuit, it was thus decided that the first order Nicol-Fleming model be used for simulating the different adsorption circuits. This model fitted the adsorption data well and the relative ease of use of this model and the fact that it is widely used in modelling adsorption circuits worldwide, it was decided that it was sufficient to arrive at the overall goal of this project. It is, however, proposed that additional work be performed on platinum, palladium adsorption models that could allow for accurate process modelling and design of such adsorption circuits for use in the precious metal industry.